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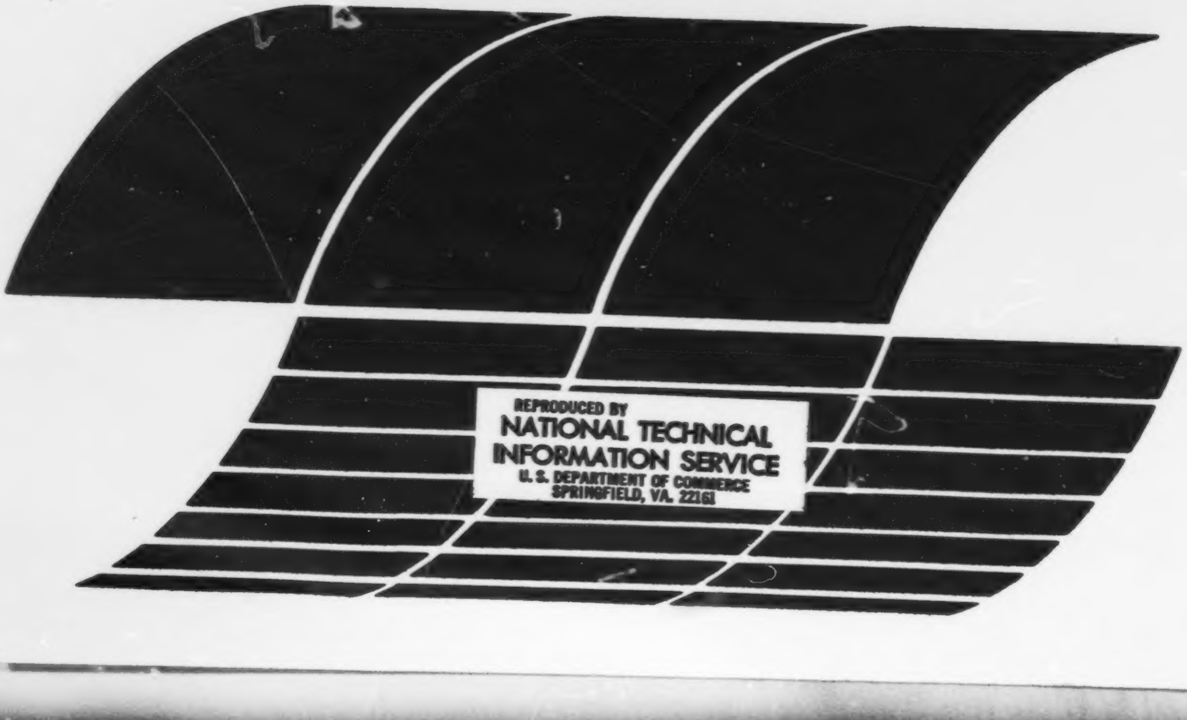

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Pollution	Pollution Control	13B
Coal Gasification	Stationary Sources	13H
Coal Preparation	Coal Conversion	13I
Regulations	Liquid Effluents	05D
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Coal Conversion Control Technology
Volume I. Environmental Regulations; Liquid Effluents

by

L.E. Bostwick, M.R. Smith, D.O. Moore, and D.K. Webber

Pullman Kellogg
16200 Park Row, Industrial Park Ten
Houston, Texas 77084

Contract No. 68-02-2198
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EPA Project Officer: Robert A. McAllister

Industrial Environmental Research Laboratory
Office of Environmental Engineering and Technology
Research Triangle Park, NC 27711

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ABSTRACT

Information has been gathered on coal conversion process streams. Available and developing control technology has been evaluated in view of the requirements of present and proposed federal, state, regional and international environmental standards. The study indicates that it appears possible to evolve technology to reduce each of the components of each process stream to an environmentally acceptable level. The conclusion has also been reached that such an approach would be costly and difficult of execution.

Because all coal conversion processes are net users of water, liquid effluents need be treated only for recycling within the process, thus achieving essentially "zero discharge." Further, with available technology gaseous emissions can be controlled to meet present environmental standards, particulates can be controlled or eliminated and disposal of solid wastes can be managed to avoid deleterious environmental effects.

Volume I focuses on environmental regulations for gaseous, liquid, and solid wastes, and the control technology for liquid effluents. Volume II deals with the control technology of gaseous emissions and solid wastes.

Volume III includes a program for economic analysis of control technology and includes the appendix.

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SECTION 1

INTRODUCTION

Proposals for development of America's natural resources to help satisfy America's energy needs invariably give coal high priority consideration. Unfortunately, coal is by no means a direct and satisfactory replacement for oil and natural gas, but conversion of coal into clean synthetic liquid or gaseous fuels promises to solve most of the problems of end use in industrial processes. This promise has spurred the development of numerous processes for production of synthetic fuels. A few of them have reached commercial status while the rest are in various stages of development in laboratories, pilot plants and demonstration plants.

The primary advantage of synthetic fuels is the transfer of the environmental problems that are associated with direct use of coal from the individual, and often small, end users to the conversion processes. Further, control technology for conversion processes may differ considerably from control technology for conventional combustion.

The objective of synthetic fuels development is to maintain and improve the quality of life through supply of energy from our natural resources without unacceptable deterioration of the environment. The Environmental Protection Agency (EPA) is responsible for the assessment of environmental factors of energy technologies and for aid in the development of controls to protect the environment. EPA has adopted a rational approach in following the development of energy systems that begins with a low level of environmental concern during the bench scale phase of the process investigation and continues with increasing awareness to realization of a comprehensive program during pilot plant and larger operations. Control technology development thus keeps pace with conversion process development.

The Fuel Process Branch of EPA's Industrial Environmental Research Laboratory at Research Triangle Park, North Carolina, is responsible for the environmental factors in the production and utilization of synthetic fuels from coal. EPA has projected the course of the fuel conversion industry and has planned environmental programs through earlier contractual arrangements that sponsored the most progressive environmental research in the synthetic fuels area and that used the available data base to indicate areas of the synthetic fuel industry that require further study. The broad survey considered all applications of the synthetic fuels technology and all proposed research and projects in control technology to ensure consideration of all predictable environmental impacts and to group together those areas of environmental importance common to a number of processes.

The EPA synthetic fuels program that is now in operation is directed toward environmental assessment and control technology development in low Btu gasification, high Btu gasification and liquefaction, including long-term contracts that emphasize data acquisition from fundamental studies, supported by research grants, and test programs at commercial facilities. Subprograms in synthetic fuels research are coordinated in the Fuel Process Branch with related programs in physical and chemical coal cleaning and studies of fuel contaminants. Coordination within the Energy Assessment and Control Division, of which the Fuel Process Branch is a part, is maintained in the area of conventional and advanced coal combustion systems.

In the course of the Pullman Kellogg study, available information concerning quantity and composition of the various emission, effluent and waste streams from coal conversion processes was gathered by literature searches and by contacts with conversion process operators. The study defines in as much detail as possible the problems that must be solved if conversion processes are to operate successfully without unacceptable deterioration of the environment.

To apply emissions control technology efficiently, goals must be set for the pollutant residuals. A major part of the total effort in the Pullman Kellogg program was the gathering and synopsizing of the present and proposed environmental regulations and standards for federal, state, regional and international jurisdictions. A summary of the most stringent of these regulations was developed for us as a standard for comparing the efficiencies of emissions control processes on the premise that a conversion plant with emissions equal to or lower than the most stringent standards could be built anywhere in the United States, Mexico,or Canada.

With problems scoped and objectives defined, information and data were gathered on available and developing control technologies. The goal was to define in as much detail as possible the effectiveness and costs of controls that may be applied to conversion process streams so that the final streams leaving the process site will meet environmental standards.

SECTION 2

MANAGEMENT SUMMARY

DEFINITION OF THE PROBLEMS

Information and data on the liquid effluents, gaseous emissions and solid wastes from coal conversion processes were gathered and evaluated from literature surveys, from communications with operators of conversion processes at bench scale, process demonstration units and pilot plants, from licensors of commercial conversion processes, from conceptual engineering designs of full scale conversion plants and from Environmental Protection Agency (EPA) and Department of Energy (DOE) contractors. Of the many conversion processes that were in various stages of development, few were found to be evaluating the composition and quantity of the effluent, emission and waste streams: the emphasis was primarily on operation of the process and data gathering on the outputs was, apparently, a secondary consideration.

Since real data on many of the conversion process output streams were lacking, incomplete or judged to be of doubtful value in attempting to scale up to full commercial operation, three process types were chosen to represent coal conversion processes:

- o Low temperature gasification, where the maximum reactor temperature is below the fusion temperature of the ash and the reactor offgas contains phenols, oils and tars. The Lurgi process was chosen for this category. There was available a considerable body of data and information on commercial operation and the composition of the output streams. The liquid effluents, in particular, constitute a type of "worst case" for application of control technology. Therefore the conclusion might be drawn that, if control technology could be successfully applied to Lurgi gasification liquid effluents, the principles of the technology would probably be successful in application to control of liquid effluents from any of the low temperature gasification processes.

- High temperature gasification, where the maximum reactor temperature is above the ash fusion temperature, the ash is discharged as a molten slag and the reactor offgases contain little or no phenols, oils and tars. The Bi-Gas process was chosen for this category. The data and information available on process effluents, emissions and wastes were gathered from literature and from a conceptual engineering design executed by C. F. Braun Company. These data were supplemented with those from commercial operation of the Koppers-Totzek process.

- Liquefaction processes, which operate at temperatures lower than the low temperature gasification processes and which produce phenols, oils and tars. The H-Coal and SRC II processes were chosen as being representative of liquefaction processes. Data and information on the emissions, effluents and wastes were drawn from the literature, from the SRC II conceptual engineering design assembled by the Ralph M. Parsons Company and from other contractors.

Tabulations and references for data found on emissions, effluents and wastes for other processes are included in the text. These data are used, in some cases, to supplement the available information on the three types of conversion processes. Included in the discussion of the total body of data and on the application of control technology are descriptions of data gaps and data variability together with discussions of problems in sampling and analysis of samples that have made difficult the scaling up of data to represent operation of full scale commercial processes.

ESTABLISHMENT OF OBJECTIVES: ENVIRONMENTAL STANDARDS

Present and currently proposed environmental restrictions relevant to contaminants in the effluents, emissions and wastes from coal conversion processes were assembled to serve as the measurement standard in evaluating available and developing control technology for such processes. The environmental restrictions in Federal and state rules and regulations were reported together with selected regional, Canadian and Mexican regulations.

The prime objective of the survey was to assemble a single source reference document of applicable environmental regulations for use in considering both present control technology capabilities and necessary future technologies for controlling pollutants from the conversion of coal to gaseous or liquid fuel.

A second objective was to summarize the most stringent of the environmental regulations so that a single source of environmental requirements representing the most restrictive of present and proposed regulations would be available. A coal conversion facility built to meet the requirements in this most stringent summary would, by definition, meet the requirements of any individual state, region or bordering country. The summary was by necessity limited primarily to those regulations of a quantitative (numerical) nature and did not include ordinances below the state jurisdictional level, since these were beyond the scope of the project. Special requirements introduced by individual states' permitting authorities were also beyond the scope of this project and were not included.

A third major objective was to provide an in-depth survey of the regulations of the selected states which had not been available previously to the extent presented in the survey. An example of the wide coverage of this survey is the inclusions of the U. S. EPA regulations applicable to Fluid Catalytic Cracking Units, Petroleum Refining category, upon reasoning that giving a broad definition to Petroleum Refining, as some states do, makes such regulations potentially relevant to expected further on-site processing of coal liquefaction products.

The first phase of the survey was concerned with Federal and state environmental regulations. As such regulations are continually being amended they can only be reported current as of a given cut-off date. The cut-off date for the Federal and state material in this report was 31 October 1977. The second phase supplements the first with a survey of regional and international regulations. Cut-off date for the second phase was 15 April 1978.

On the premise that the survey should be as broad as possible, it was decided that expanding the material considered relevant would be preferable to restricting it. Consequently, whenever it appeared that a particular standard or regulation might have at least some present or potential relevance, it was included in the survey. This approach was also advantageous with respect to use of the survey by project personnel as a source of guidelines to demonstrate the type and degree of restrictions placed on environmental contaminants.

The coverage of the survey was intentionally made as broad as possible to present the widest and most divergent restrictions in effect at both the Federal and state levels. As the commercial coal conversion facilities which are the underlying subject matter of this project are all yet to be built, only regulations pertaining to new facilities, as opposed to existing facilities, were considered and included.

Selection of the states to be included in the survey was based on the reported availability of coal deposits within the states, since economic factors favor sites near coal deposits for possible coal conversion plant locations. Accordingly, the environmental laws, regulations and standards for the following 22 states were included with the federal restrictions:

Alabama
Alaska
Colorado
Idaho
Illinois
Indiana
Kansas
Kentucky
Missouri
Montana
New Mexico
North Dakota
Ohio
Oklahoma
Pennsylvania
Tennessee
Texas
Utah
Virginia
Washington
West Virginia
Wyoming

The requirements as established by the U. S. Public Health Service Drinking Water Standards, 1962, and the Interim Primary Drinking Water Regulations were synopsized and included in the survey together with a review of standards and guidelines established by the Delaware River Basin Commission, since the authority of this regional commission extends over geographical, rather than political, areas and therefore considers the area environment unconfined by artificial boundaries. It was found that the standards and guidelines adopted by the Susquehanna River Basin Commission are those of the states affected by the Commission.

Further consideration of the argument that environmental effects are not limited by political boundaries led to the inclusion in the survey of the standards and guidelines that have been established by Mexico and Canada. The Mexican regulations are federal actions, while in Canada both the Dominion and the provincial governments have enacted standards and guidelines. Therefore, Mexican federal standards, Dominion of Canada standards and guidelines, and the standards and guidelines of the provinces of Alberta and British Columbia became part of the survey. The two provinces were chosen because their boundaries are continguous with those of Montana, Idaho and Washington, where much of the U. S. western coal reserves are located.

Finally, the rules and guidelines established by U. S.-Canadian International Joint Commission were included in the survey, since these are primarily concerned with the Great Lakes and the St. Lawrence River areas and thus complete the regulatory coverage of the northern U. S. border.

The listings of the most stringent environmental standards and guidelines were compared to the Multimedia Environmental Goals (MEGs) wherever possible. The MEGs are being developed by EPA as estimates of desirable ambient and emission levels of control and, as such, are an integral part of EPA's environmental assessment approach. Concentration categories for contaminant substances in water effluents, water ambient bodies, air effluents and air ambient bodies were compared wherever numerical information was available. Comparisons were developed for 43 substances. Provisions of other regulations, although relevant and important to coal conversion, were stated only in terms of allowable rates, were thus not comparable to MEG concentration data and were not included in the comparison. Solid waste media comparisons could not be made because few of the solid waste synopses contained numerical standards that could be used for comparison.

The value of the comparison of the most stringent air and water environmental standards with MEGs lies in the establishment of a basis for recommendations for future environmental goals.

LIQUID EFFLUENT TREATMENT

In discussions with water treating process licensors and equipment vendors the listing of most stringent standards for water quality was presented as the target for pollutant residuals. Most licensors and vendors felt that the most stringent standards were attainable, but none could make firm statements without experimentation on actual, representative samples. All the licensors and vendors appeared desirous of participating in some arrangement, such as a contract with DOE or a subcontract with a DOE prime contractor, in which samples could be tested or the performance of their rental treatment units in pilot or demonstration plants could be monitored.

All the licensors and vendors agreed that treatment of liquid effluents for recycling to the conversion process unit as process water would be easier and cheaper than treating the individual streams to meet the standards for discharge to receiving waters: specifications for reuse of water as cooling tower makeup and boiler feedwater makeup are much less stringent than drinking water standards. The general opinion was that treatment methods for recycling would be the same as, or similar to, commercial treatment procedures, but that experimentation would be required for confirmation.

Based on the statements and opinions of the treatment process licensors and equipment vendors, supplemented by Pullman Kellogg commercial experience in the field, tentative water treatment flowsheets were assembled in which maximum reuse of water was projected. Water quantities for the flowsheets were obtained from conceptual designs published by DOE sponsored contractors C.F. Braun (Lurgi gasification, producing phenols, oils,and tars, and Bi-Gas gasification, producing no phenols, oils,and tars) and Ralph M. Parsons (SRC II liquefaction). In these flowsheets no discharge of water would be practiced; water losses would be from cooling tower evaporation, the water associated with quenched ash or slag and the water in wet inorganic sludges from water treatment. Provision would be included for collecting leachate and runnoff from the solid waste disposal area and returning the liquid to the conversion plant for ash quenching or treatment for recycling. These provisions in effect close the loop and promote the concept of "zero discharge." It should be noted that the term "zero discharge" only describes the effect of avoiding discharge of treated water to receiving bodies and does not imply that there is no liquid leaving the conversion plant.

Commercial methods for treatment of raw water and wastewater were investigated in detail for capability, efficiency, limitations, case histories, wastes produced from the treatment process, possible problems and possible improvements. Methods included:

- Chemical Precipitation (Softening)
- Reverse Osmosis
- Ion Exchange
- Evaporation
- Electrodialysis
- Oil Separation
- Phenol Extraction
- Stripping and Ammonia Recovery

- Chemical Coagulation and Flocculation
- Flotation
- Biological Oxidation
- Filtration of Biological Oxidation Effluent
- Biological Oxidation Sludge Handling
- Carbon Adsorption
- Chemical Oxidation (chlorine, ozone, hydrogen peroxide, etc.)

Capital and operating costs for these processes are often stated in the literature as "typical" or "classical." It was found that use of these stated costs in economic evaluations could lead to gross errors because these costs may be in error due to inflation, incomplete data,and unclear or confusing presentation. Correspondence and discussion with treatment process licensors and equipment vendors was found to be most productive of reliable costs, based on Pullman Kellogg's best estimates of quantities and compositions of the feed streams to treatment and desired contaminent levels in the treated streams. Process licensors in general furnished complete information on equipment costs, installed costs and operating costs, while equipment vendors usually furnished only cost of equipment "knocked down," f.o.b. factory.

Data on various methods of measuring the effect of inflation or equipment and construction costs were assembled into curves for use in updating costs reported in the literature. Factors were developed for estimation of total capital cost when only the vendor's equipment cost is available. With use of the curves and factors the published capital costs of treatment processes proposed by, among others, Water Purification Associates, Bechtel and Associated Water and Air Resources (AWARE) and for plants installed by Pullman Kellogg, were tabulated and compared.

A Pullman Kellogg budget estimate for a biological oxidation system, preceded by flash mixing, flocculation, and dissolved air flotation, was developed with the aid of Envirotech Process Equipment personnel in Houston and Salt Lake City. Feed to this system had previously undergone (by calculation) oil separation, phenol extraction, stripping,and ammonia recovery.

Good, recently-published cost data were gathered for side stream softening of cooling tower blowdown, oil separation and regeneration costs of granular activated carbon. Also included were costs for a system installed by Pullman Kellogg for a Peoples Gas SNG plant that included ion exchange, evaporation,and spray drying of final solids for disposal.

Operating costs of water treatment processes were explored as far as time permitted.

Best data were those received from process licensors such as American Lurgi (Phenolsolvan), Chevron Research and U.S. Steel (Stripping and Ammonia Recovery),and Zimpro (Wet Air Oxidation). Values used for utilities, labor, and chemicals are clearly stated so that all operating costs are on a common basis. Zimpro furnished operating costs for biological oxidation using powdered activated carbon and wet air oxidation of circulating carbon-sludge. Partial operating costs were tabulated for flotation, biological oxidation, evaporation, demineralization, and reverse osmosis, as supplied by equipment vendors.

Budget cost estimates for treatment processes were received from the following licensors and vendors:

- Stripping and Ammonia Recovery:
  Chevron Research WWT Process (2 cases)

- Stripping and Ammonia Recovery:
  U. S. Steel Phosam-W Process (2 cases)

- Biological Oxidation with Powdered Carbon and Wet Air Oxidation Regeneration:
  Zimpro, Inc. Wastewater Reclamation System (2 cases)

- Wet Air Oxidation of Raw Wastewater:
  Zimpro, Inc. WAO Process

- Disposal of Biological Oxidation Sludge:
  Zimpro, Inc. LPO (Low Pressure Oxidation) system for thermal conditioning of sludge.

- High Purity Oxygen Activated Sludge:
  Union Carbide UNOX process (2 cases)

- Phenol Extraction:
  American Lurgi Phenosolvan process

- Raw Water Treatment Process:
  L*A/Water Treatment (division of Chromalloy)
  Lime softeners, gravity filters, clear wells, Zeolite softener, demineralizers, condensate polisher, BFW deaerators, and reverse osmosis preceded by pressure filtration.

- Evaporator-Crystallizers:
  Goslin division of Evirotech, Inc.
  Forced circulation, six-effect evaporator-crystallizers

Integrated wastewater flowsheets were assembled for the three coal conversion processes(Lurgi and Bi-Gas gasification and SRC II liquefaction) with utilization of developed information on the efficiencies of wastewater treatment processes. Sour water analyses were selected from the available data sources that were considered to be representative: Lurgi for low temperature gasification, Koppers-Totzek for high temperature gasification, and H-Coal for liquefaction. The influent and effluent analyses for the several water streams were evolved from best engineering judgement of in-house experts, licensors and vendors, who considered them generally reasonable in lieu of actual treatment tests.

Capital cost for the Lurgi wastewater treatment scheme was estimated at $42,000,000. An alternate case that included reverse osmosis showed a small saving in capital cost, but substantial operating cost savings. Several alternatives were considered as were the effects of possible additions which might be necessary, particularly for coals of high chlorine content. Many possible alternate methods for organics removal were suggested, such as anaerobic digestion, wet air oxidation, various schemes for powdered activated carbon use in biological oxidation and side stream cooling tower blowdown variations.

For the Bi-Gas flowsheet, the wastewater treating scheme is simpler and capital cost is estimated at about $23,000,000. Some alternates and possible problems or additions are discussed as a critique of this integrated system.

Two versions of wastewater treating were compared for the liquefaction flowsheet. The first was the simple low cost scheme used by Parsons in their conceptual design, where stripped wastewater was recycled directly to the waste heat boiler system for the process (Bi-Gas) gasifier, thus destroying any phenols, oils and tars. Only raw water was treated, and 3,000 GPM of treated wastewater was discharged to the river.

The second scheme was a more conservative, higher cost system for zero discharge which is recommended by Pullman Kellogg.

Parsons water treating capital cost, exclusive of that incurred in the waste heat boiler system, was $26,220,000. Pullman Kellogg's wastewater treating system capital cost was estimated at $33,500,000, not including raw water treating. If raw water treating costs were included, the total cost for the Pullman Kellogg scheme became $49,270,000 compared to $26,220,000 for the Parsons scheme which did not include the unknown costs due to injection of wastewater into the waste heat boiler system.

Partial operating costs for the integrated schemes, including the major units, were developed, ranging from $17,000/day for liquefaction to $31,000/day for Lurgi gasification, for wastewater treatment only, not including raw water treating.

Water treating technology which is under development or not widely used at present was listed and references noted. Time did not permit pursuit of efficiency or cost data on any of these.

GASEOUS EMISSION TREATMENT

Gaseous emissions include entrained particulates. Technology for control of these gases and solids involves in most cases application of a single control process or a single assembly of control equipment, in contrast to the systems approach that is needed in water treatment.

The list of most stringent standards was used as the criterion for selection of control technology and evaluation of efficiencies of the various alternatives. Discussions with control process licensors and equipment vendors supplemented information gathered from literature searches and in-house information.

Essential information on emissions from recovery of products and byproducts in coal conversion plants was to be supplied to Pullman Kellogg as their starting point for efforts on application of control technology of such efficiency that the final gaseous streams reaching the atmosphere would be of a quality equal to or better than the most stringent of environmental standards. Because this information was not supplied, Pullman Kellogg were forced to develop process flow sheets and material balances from published information in order to estimate compositions and quantities of gas streams from recovery processes. The time required for this effort shortened the time available for investigation of control technology and therefore the investigations take the form of general descriptions of control technology, examples of application of the technology, evaluation, wherever possible, of means of increasing control process efficiency, and cost information.

The Lurgi Dry Ash process was selected as the base gasification case for study of integrated schemes for emissions control. The flow diagram and material balance, assembled from the conceptual designs of C. F. Braun, Cameron Engineers and Pullman Kellogg for operation of the Lurgi process on western, low sulfur coal, were primarily directed toward establishment of the operating characteristics of the sulfur recovery unit, the composition of the offgas stream from sulfur recovery and the required operating characteristics of the unit for control of the sulfur emissions from the coal conversion plant. As an alternate case, a flow diagram and material balance were developed for eastern, high sulfur coal.

The Lurgi Dry Ash process produces phenols, oils and tars that are separated from the gas stream and either processed further for sale or sent to an incinerator/boiler. Because of the quantity of materials that must be disposed of by incineration, the incinerator/boiler is an important part of the coal conversion plant, its operation is closely integrated with the sulfur recovery unit and treatment of its offgases provides a second source of product sulfur.

On the other hand, the high temperature gasification processes, exemplified by the Bi-Gas process, produce little or no phenols, oils and tars and the importance of the incinerator/boiler as a means of waste disposal is reduced. Process steam is raised more from coal and less from waste, the overall sulfur balance changes and the demands on the sulfur recovery and emissions control units change. The overall sulfur balance for Bi-Gas operation was based on the C. F. Braun conceptual design for operation with low sulfur coal and was recalculated to demonstrate the changes with use of high sulfur coal.

For liquefaction, the Ralph M. Parsons conceptual design for the SRC II process was selected as representative. As in the gasification processes, calculation of the material balances was primarily directed toward determining the operating characteristics of the sulfur recovery unit, the offgas stream composition and the demands on the unit for control of the sulfur emissions.

Control technology proposed by others in the conceptual designs and pertinent reports was reviewed and compared to the most stringent environmental standards, with the conclusion that most of the proposed techniques and methods would not meet the most stringent standards.

The commercial emission control methods that were considered to be most important for application to coal conversion processes, and that were investigated, are:

- Processes and techniques for control of nitrogen oxides
- Processes and techniques for control of sulfur dioxide
- Processes for control of hydrogen sulfide
- Techniques for control of particulates
- Control of cooling tower drift
- Other control techniques applied to hydrocarbons, lock hopper vent gas, ammonia, ash quench vent gas, and miscellaneous leaks.

Capability, efficiency, limitations, case histories, wastes produced, costs, possible problems, and possible improvements were determined for the commercial processes that may be applied to emission control.

Nitrogen Oxides Control

In Lurgi gasification of both eastern and western coal the waste gas streams and the liquid byproduct and waste streams have a heating value equivalent to about 58 percent of the total plant energy requirement and may be fed to an incinerator/boiler with coal providing the remainder. Either boiler modifications or liquid fuel denitrogenation were shown to reduce emissions sufficiently to meet present environmental standards. Both may be applied if further reduction is required. If still further reduction is needed, there are flue gas denitrogenation processes, both dry and wet, that may be applied.

With Bi-Gas gasification, combustible liquid byproduct and waste streams are eliminated, the waste gas streams are reduced drastically and the steam generator operates primarily as a coal-fed boiler. Boiler modifications appear to be sufficient to reduce nitrogen oxides to meet present most stringent environmental standards.

In liquefaction, the conceptual design uses a slagging gasifier to produce fuel gas for boiler fuel. With staged combustion and low excess air the expected nitrogen oxides emissions are estimated to be less than the most stringent standards.

Sulfur Dioxide Control

Sulfur in the feed coal may be reduced physically, chemically, or with a combination of both. Sulfur dioxide formation in the boiler may be reduced by fluidized bed combustion. Sulfur dioxide in the flue gases may be reduced by scrubbing or by dry absorption or reaction. Combinations of these may be selected by economic analysis and consideration of the marketability of recovered sulfur.

For this study, the western and eastern coals were assumed to be fed without prior treatment. Sulfur was recovered from conversion process streams by the Claus process followed by the Beavon offgas treatment process. Sulfur dioxide was removed from incinerator/boiler flue gases by the citrate process which used part of the Claus process feed as a source of hydrogen sulfide and recovered more sulfur. Final process offgases would meet present most stringent environmental standards.

Hydrogen Sulfide Control

In the acid gas separation step of the conversion processes the carbon dioxide stream, contaminated with hydrogen sulfude, was sent to the incinerator/boiler, while the hydrogen sulfide stream, joined by streams from sour water stripping, was fed to the Claus process and to the flue gas desulfurization scrubber. Offgas from the Claus process was treated in a Beavon process unit to reduce the level of sulfur compounds in the final offgas low enough to meet the most stringent environmental standards.

Splitting the hydrogen sulfide stream reduces the required size of the Claus and Beavon units, supplies the citrate process unit with its needed hydrogen sulfide and increases the overall marketable sulfur yield.

Particulate Control

Assuming that 80 percent of the ash content of the coal fed to the incinerator/boiler became flyash, then a combination of cyclones and electrostatic precipitators was shown to reduce particulate emissions to levels well below the most stringent standards. It was noted, but not evaluated, that the spray coolers that precede the flue gas desulfurization scrubber will remove nearly all particulates remaining after the cyclones and the electrostatic precipitator may possibly be eliminated.

Control of Cooling Tower Drift

Droplets of cooling tower water, entrained in the air flowing through the tower, may be hazardous, toxic, or a nuisance due to the content of dissolved or suspended chemical compounds, particularly when the cooling tower basin is used as a catchall for boiler blowdown and miscellaneous waste waters. Commercially proved designs of drift elimination systems have been developed that reduce by over 90 percent the drift that is normally experienced with conventional two-pass eliminators.

Other Control Techniques

Calculation of possible emissions of hydrocarbons, carbon monoxide, and ammonia from the incinerator/boilers of coal conversion processes indicated that these emissions were well below the most stringent environmental standards.

Coal lock hopper vent gas cannot be sent directly to the atmosphere because of its high content of hydrocarbons and sulfur species. The recommended procedure involves bleeding the pressurizing gas from the empty lock hopper into the fuel system or the reactor gas system, displacing residual gas in the hopper with carbon dioxide or nitrogen and directing the purge stream to incineration.

Vent gases from ash quenching are principally steam with entrained ash particles, but some hydrocarbons may be formed from organic materials in the quench water and unreacted carbon in the ash. The recommended sequence is separation of solids in a wet cyclone, scrubbing by direct contact with water and incineration of the noncondensables.

Emissions from storage vessels may be controlled by incineration, by refrigeration to condense vapors, by scrubbing systems that use low volatility solvents or by adsorption systems. Choice of method depends on type and quantity of emission and, finally, on economics of the control systems.

Costs of Control Techniques

Processes selected for investigation were considered to be representative of the best available technology. The developed costs are simplified hypothetical cases, intended to demonstrate the types of studies that would be required for more rigorous treatment, and cannot be interpreted or used as definitive estimates. In-depth studies would be needed in order to make specific process recommendations.

Capital costs and, wherever possible, operating costs were developed as follows:

- Particulate control: installed cost vs. capacity for cyclones, wet scrubbers, fabric filters,and electrostatic precipitators. No operating costs.

- Coal desulfurization: capital and operating costs for processing 25,000 tons per day of coal through a heavy media plant and through the Meyers chemical cleaning process were developed. The heavy media process showed lower capital and operating costs than the Meyers process but removed only 75 to 85 percent of the pyritic sulfur in contrast to the 95 percent removal in the Meyers process. In terms of sulfur removal cost, then, the two processes are nearly equivalent.

- Sulfur dioxide control during combustion: capital and operating costs of fluidized bed combustion (FBC) of coal with capture of sulfur dioxide by limestone was contrasted with those for combustion in a conventional boiler followed by flue gas desulfurization. Cost of operating with two coals at two capacities in a single boiler installed in a coal fired power plant, in a single boiler installed in an oil fired power plant and in a grassroots boiler plants with backup were developed. Where no sulfur dioxide controls are needed, the conventional boiler has lower costs than the FBC. For high sulfur coals, the FBC appears to be the better choice.

- Flue gas desulfurization: capital and operating costs were developed for limestone slurry, lime slurry, magnesia slurry,and catalytic oxidation processes. Capital investments for the three throwaway processes are lower than the product recovery process, with lime slurry being the lowest. Operating costs of magnesia slurry are highest, followed by catalytic oxidation and lime slurry, with limestone slurry the lowest. In a comparison of limestone slurry and the citrate process, the citrate process with credit for sulfur sales is a close competitor of limestone slurry.

- Sulfur dioxide control by citrate process FGD alone or with pyritic sulfur removal plus FGD: the comparison of operating costs of FGD alone and of removal of 80 percent of pyritic sulfur by heavy media washing followed by FGD as required to meet most stringent standards showed that costs for FGD alone were about 60 percent of those for coal cleaning plus FGD. A similar comparison for oil firing showed that hydrodesulfurization of the feedstock is more economical than FGD.

- Nitrogen oxides control: operating costs for flue gas denitrification alone, fuel hydrotreating alone and combined 100 percent hydrotreating plus flue gas denitrification were developed for control of nitrogen oxides from combustion of tar and oil wastes from coal conversion. Flue gas denitrification alone was shown to be far superior to the other methods, with fuel hydrotreating alone next highest and the combustion method costing nearly three times flue gas denitrification alone.

- Combined sulfur dioxide and nitrogen oxide control by fuel oil hydrotreating: capital investments and operating costs for gas oil and residual oil hydrotreating to remove 92 to 97 percent of the sulfur and 80 to 90 percent of the nitrogen were developed. Operating costs for hydrotreating alone, hydrotreating most of the oil to meet sulfur dioxide standards then treating the flue gas with the UOP/Shell process to remove nitrogen oxides, and using the UOP/Shell process to remove both oxides from the flue gas with prior hydrotreating were calculated.

To meet the nitrogen oxide goal of 150 ppm per million Btu, hydrotreating alone is superior. If the goal is lowered to 90 ppm, the combined processes are superior, although it is possible that boiler modifications and hydrotreating would suffice. The case without hydrotreating does not appear to be competitive.

- Combined sulfur dioxide and nitrogen oxide control for coal fired boilers: flue gas treatment by removal of sulfur by the citrate process with nitrogen removal by the UOP/Shell process was contrasted with simultaneous removal of both oxides with the UOP/Shell process. To meet the most stringent standards for both oxides the first case shows a lower operating cost.

- Hydrogen sulfide control: capital and operating costs for the Claus-Beavon combined recovery and offgas treatment were developed as functions of capacity. Time did not permit full evaluation of the effect of variables and costs and the economics of the various process combinations.

SOLID WASTE CONTROL

Fugitive Dusts

Fugitive dusts are generated when a stream of dry solids falls freely in air as at a conveyor transfer or discharge point, when the stream of solids is agitated as at railcar dumping or reclaiming from pile storage and when wind blows across piles of solids. There is little information on fugitive dust generation and therefore it was necessary to attempt to define the problem by a series of compilations and extrapolations of coal crushing data and other applicable information. From this study the conclusion was reached that on the order of 0.5 to 4 percent of the coal feed to a conversion plant could reach the ground 500 feet from the storage pile with a 20 MPH crosswind, or 75 to 700 tons per day of coal feed to a 250 billion Btu per day gasification plant.

Water sprays, particularly when a wetting agent is included, applied at rates of about 2 gallons per ton, are fairly effective in suppressing dust as it is generated at car dumping stations, transfer points, stackers and reclaimers, and for a short time in suppressing fugitive dust evolution from the storage pile. Use of larger quantities of water in flooding type sprays may help to alleviate the immediate problem but this does not solve the long-term storage problem.

Total wetting of the coal as it is received is suggested as a means of increasing the efficiency of dust suppression. From 1,300 to 3,900 gallons of water are estimated to be required for flooding a 100 ton car of coal, depending on the coal type.

Chemical binders for use on coal in "dead storage" piles and also for use on coal during rail transport were evaluated for efficiency and cost. Physical binders, such as asphalt or road tar, and compaction were investigated as means of suppressing dust.

The concept of dust elimination--instead of suppression--is advanced as a means of dealing with the dust problem by curing the cause, rather than treating the symptoms. Dust particles less than 500 micrometers in diameter are separated from the larger coal particles at or near the point of creation of the dust by dry or wet means then these particles are agglomerated into masses that are 500 micrometers in diameter and larger.

Agglomeration may be by compaction, by briquetting or by granulation in a pan or drum granulator. Binders will probably be needed, such as tars and oils from coal conversion, asphalts, tars or waxes from petroleum refining, bentonite clay, or starch. Evaluation of the agglomeration means is needed to equate efficiency with capital and operating costs. Dust elimination according to this concept eliminates most or all of the dust suppression methods and equipment, with their associated operating costs.

Collection and disposal of coal dust is discussed, together with means for control of such other process dusts as ash/slag, limestone and spent catalysts.

Cost are developed for dust suppression by water sprays and chemical binders. Insufficient data and information were available on methods of dust elimination to evaluate their economics and efficiencies.

Solid Waste Disposal and Management

Solid inorganic wastes from coal conversion plants consist of ash or slag from the gasification reactors, ash or slag from the boilers or incinerator/boilers, FGD sludge if the process is used, inorganic salts in solution and suspension and spent catalysts, of which the ash/slag composes on the order of 99 percent. For this study the organic wastes from biological treatment of wastewaters were assumed to be sent to the incinerator or to the gasifier reactors. Depending on coal type and conversion process operating characteristics, the total solids volume was shown to range from 10.4 to 25 million cubic feet per year.

The overall water treatment schemes that were developed in this study required recycling as much water as possible or practical and minimizing or concentrating the wastewaters that could not be recycled but which carried inorganic salts. This minimum stream was finally combined with dewatered quenched ash into a mixture containing 70 to 80 percent solids: damp enough to be dust free and dry enough to be handled in trucks, on conveyors or by aerial tramway to the disposal area. These methods of transport were evaluated for conversion plant solids by extension of information on transport of chemically stabilized FGD sludge, with the knowledge that the conversion plant solids present fewer and less severe problems than do FGD sludges. Means of increasing the solids content of the waste stream by oil-fluidized evaporation of inorganic sludges were investigated and found to be worth considering.

The economics of construction of solids disposal areas were investigated to demonstrate that the dimensions of pits constructed partly above grade and partly below grade can be optimized with regard to plan area occupied and interior surface area.

Interposition of an impervious membrane between the solids and the environment offers an attractive means of avoiding the environmental damage that could be caused by leachates from the disposal area reaching ground or subsurface waters. The performance and installation costs of a number of membranes were evaluated through extension of data that were originally developed for sanitary waste disposal.

Chemical stabilization of the wastes will reduce the permeability of the mass to a level comparable to that of hard packed clay. The operation and economics of three commercial systems were evaluated as an alternative to the installation of a membrane to prevent leaching.

Recognizing that leaching will take place, means were evaluated for collecting leachate for return to the conversion process plant for reuse as ash quench water or, following appropriate treatment, for recycling to process.

ECONOMIC ANALYSIS AND PROGRAM EMPHASIS

Existing and developing control technology was examined to determine applicability to the problems surrounding coal conversion. The individual control processes were examined to determine the possibilities for increasing efficiency, the probability that the process would be used, the availability of adequate information to permit economic evaluation and the availability of alternate control technology, then a list of candidate processes was assembled for economic evaluation.

A program was proposed for economic evaluation of those processes selected from the list of candidates. The program included assembly of cost models to reflect required control efficiency, size and location, with the intent that the cost models could be used as modules and be added to various combinations to yield the total cost of environmental control for the conversion processes.

SECTION 3

CONCLUSIONS

Lack of definitive information and data on the compositions and quantities of the liquid effluents gaseous emissions and solid wastes from operation of coal conversion processes required the development of estimates of the streams in order to assess the efficiency, operating characteristics and costs of commercially available and developing control technology. The estimates of compositions and quantities, developed from literature searches, personal communications and application of best engineering judgment, were vital to the success of the Pullman Kellogg project. The estimates cannot however, substitute for data derived from actual process operation. Consequently, the treatment schemes and the projected results of these schemes may be considered as near approximations of expected results, but cannot be considered as being definitive.

Real data and information on the effluents, emissions, and wastes from coal conversion processes are being developed through efforts in EPA programs, such as "Level 1 Environmental Assessment," and in the programs of others. These data must be compared to the estimated values used in this project to determine whether or not changes are required in the proposed control technology or in the control philosphy.

A simple compilation of Federal, state, regional, Canadian and

Mexican regulations and standards, existing and proposed, is an unwieldy tool for measurement of the efficiency of control technology as it is applied to the conversion process streams. By synopsizing the regulations and then, as far as possible, listing the most stringent of the regulations for each potential pollutant, a measurement standard is established such that a conversion process plant in which the effluent, emission, and waste streams are controlled to meet the standards in the master list will meet standards anywhere in the U. S., Canada, and Mexico.

EPA's programmed development of "Multimedia Environmental Goals" must be included in future evaluations of control technology versus environmental impact, and wherever bases for possible future environmental standards or goals are considered.

Technology appears to exist for control of the components of most of the effluent, emission, and waste streams from coal conversion plants, according to the best estimates of control technology licensors and vendors of control equipment, such that treatment of the streams for release to receiving bodies of water, in accordance with the most stringent environmental standards, might be possible. Such an approach to the control problem is, in general terms, difficult and uneconomical. Pullman Kellogg have developed a sound, practical engineering approach to the problem from the viewpoint of the conversion process operator and have concluded that, for easiest and most economical operation, control technology appears to be available to:

- Treat liquid effluent streams only enough to allow the treated water to be recycled to the conversion process: control technology required is much less severe, effluent to receiving bodies of water is drastically reduced or eliminated, treatment costs are attractively reduced and raw water usage is significantly decreased. Only the irreducible minimum of inorganic salts remain for final disposition

- Wherever possible, gather gaseous emission streams for collective treatment and application of emissions control technology and treat individual streams only when composition or quantity demands it. Recover usable byproducts, such as sulfur, to reduce control and disposal problems and expense and to increase revenues

- Suppress or eliminate dusts, particularly coal dust, from handling, transportation, storage and reclaim operations. Dispose of solid wastes by isolating them from the environment or by chemically treating them so that environmental impact, such as by leachates penetrating adjacent soil, is reduced below that allowed in the most stringent standards

The treatment methods proposed and developed in the Pullman Kellogg study are best estimates of the performance of control processes on estimated stream compositions and quantities. Sampling of conversion plant effluents, emissions,and wastes is needed to supply the licensors of control technology and the control equipment vendors with sufficient feedstock for testing in their proposed process schemes to assure that real results will meet the present environmental standards.

Development of technology so that conversion process effluents, emissions,and wastes can be controlled to meet or to be better than possible future environmental goals requires supply of adequate quantities of representative materials to technology developers.

Investigation into stream treatment processes is needed to determine the maximum practicable process efficiency in comparison to present and proposed environmental standards as an aid to defining areas for development of new control technology.

Since the streams or the components of the streams can be rendered harmless to the environment in several ways (e.g., containment, recycling, separation and destruction, conversion), choice of the treatment methods finally devolves to the efficiency and economics of the methods based on sound engineering decisions.

SECTION 4

RECOMMENDATIONS

FOR PROJECTION OF FUTURE ENVIRONMENTAL GOALS

A thorough study is needed of the health and ecological effects of possible contaminating substances. There are a number of good sources that should be utilized to gather data on criteria, including MEG related studies and charts, Threshold Limit Values (TLVs) of various organizations such as ACGIH, NIOSH studies and recommendations, OSHA regulations and reports, U. S. Public Health Service studies, guidelines and standards, NAS/NAE Water Quality Criteria, Chemical Industry Institute of Toxicology reports,and reference compilations such as "Industrial Hygiene and Toxicology," by Interscience Publishers. Such a study would also be helpful in better evaluating the current and proposed environmental regulations, since the basis on which the regulations were established is generally not known. This is one of, or possibly the highest of, priority recommendations, provided resources permit such an approach. The MEG comparison that was made in this report was a preliminary or first step in this direction and could also be expanded on as a more detailed study.

Current and proposed applicable regulations of jurisdictions other than those selected in this project should be reviewed. These could include other highly developed countries, such as

Germany, Japan, France, United Kingdom, and Sweden, other states with either newly discovered or potential coal deposits or known to have very stringent regulations, such as California or Los Angeles County, and other possible international bodies of regulations (the latter not likely to be highly fruitful).

Complete process designs of several favored conversion plant configurations should be assembled with different scenarios for various coal feeds coupled with the use of programmed modelling techniques to determine ambient concentrations of pollutants in the areas outside of facilities and at different altitudes or depths. This would allow complete analyses and comparison of different regulations regardless of their bases or units or their presentation as an equation. Ambient media and effluent regulations could be analyzed equally well.

A thorough study could be accomplished of applicable substances as carcinogens, mutagens or teratogens and the limitation levels dictated thereby. Studies of the concept of "zero threshold pollutants," as referred to in the MEG report, would be recommended here.

A "best future technology" approach could be developed, based on a study of estimates, forecasts and reports on developing technology. National and international economic considerations could also be studied and analyses made both with and without them factored in. Projected energy, fuel and transportation availability as possibly affecting such concerns as national security and utility reliability might also be considered.

With regard to establishment of future environmental goals, note should be taken that some of the regulations already in operation are more indicative of future standards than others. The New Mexico air regulations for gasification plants are probably close to five year levels and possibly somewhat beyond. British Columbia water regulations and International Joint Commission proposed water regulations are probably close to five year levels. Various Water Act mandates are very good indicators for the level of future water standards. "Fishable, swimmable, navigable waters" are mandated by July 1, 1983 and zero discharge of pollutants to navigable waters by 1985. Toxic pollutants in toxic amounts are already prohibited. A closer review of the existing and proposed laws and of the regulations presented herein is highly recommended as an aid in projection of future regulations. A review of the synopses under a different set of guidelines might produce different levels for current most stringent regulations for some substances. Discretion is necessarily involved in the most stringent regulation selection process and philosophy or guidelines used probably resulted in close to the lowest allowable levels possible from such an analysis.

Regardless of other approaches or methods pursued, a continuation of the review and updating of applicable Federal and state legislation and regulations as promulgated and published in the Federal Register and other timely periodicals or issuances of the jurisdictions and new bases for closely related regulations should also be followed for further insight. Mandates of new environmental laws as they are passed must be identified and interpreted in the light of possible effects on regulations or regulation of new parameters. Following new regulations yet to be promulgated under current legislative mandates, such as for more ambient air criteria substances and more industries under new source performance standards, also will be a necessity.

In air pollution control the effect of the prevention of significant deterioration (P.S.D.) and emission offset regulations, as these control methods mature, on point source regulations for air criteria substances will have to be taken into account.

These new regulatory concepts might have a major effect on future regulations of the point source, fugitive emissions, and ambient media type. Mandated changes in state implementation plans due to lack of progress in attaining criteria substance ambient standards must also be reviewed and analyzed for probable effects on other future regulations.

Water regulations based on use of specific supply objective requirements (as in Section VI of Canada Federal Water synopses) in various jurisdictions might be used to aid in forecasting receiving water and emission levels regulations. If brought into more general use, this type of regulation might control the setting of receiving water standards just as these generally control point source effluent concentration allowables. Drinking water standards are the best example of water use standards already commonly in effect.

Study of relevant substances in the light of elimination of discharge (EOD) type emission level goals, another concept being used in MEG studies, is needed. These goals would be the most stringent and would be based on the premise that ambient pollutant concentrations should not exceed natural background concentrations. Dilution factors are used to put ambient concentrations in terms of effluents. Rural air atmospheres and drinking water and seawater are frequently used in studies to aid in the study indication of natural background concentrations.

FOR STUDIES OF LIQUID EFFLUENT TREATMENT

Recommended Laboratory Verification of Available Treatment Schemes

Actual coal conversion process wastewaters should be supplied to licensors and vendors for testing in their own laboratories or rental treating equipment supplied by the vendors should be operated directly in the pilot plants.

For oil separation, laboratory investigations are needed to more clearly determine whether or not there are emulsion problems and, if problems are apparent, the best means to break the emulsions. Reports of pilot plant operations have not included in clearly usable form the concentrations of fly ash, char, unconverted coal fines, or other insoluble solids that may be present in the sour water. Information and data are needed from which any effects of the solids on oil separation may be ascertained and the best means to deal with the solids may be determined. Equipment vendors could be of great help in this respect.

Although there are indications that single stage steam stripping will drive off carbon dioxide and hydrogen sulfide to sufficiently low levels, and ammonia to the level needed for biological oxidation, a laboratory program of confirmation is recommended. The amount of ammonia actually required for biological oxidation must be established for use as a guide for the stripping investigations. (Our theory is that the biological pond need have no more ammonia than can be stoichiometrically used by phenols and other easily biodegradable compounds in the first stage. Cyanates apparently do not begin to degrade until these compounds are gone and when cyanates degrade they produce ammonia. Residual ammonia from the final biolgical stage will be hard to control unless long sludge age is used, probably with powdered carbon.)

The problem of obtaining low ammonia residuals when biological oxidation is not used can be solved, we believe, by two-stage stripping with lime clarification between the stages. Lime addition to pH 9.5 to 11 will be beneficial in many ways: it precipitates tars, suspended solids, and trace metals as well as freeing "fixed" ammonia from ammonium salts of acids, such as ammonium chloride. Early simulation of this method on process condensate derived from coal conversion processes operating at high temperature and containing no phenols, oils, or tars is recommended.

In flotation it is recommended that pH adjustment before flotation with carbon dioxide and with sulfuric acid be compared economically and operationally by experiment, including additives to obtain best oil separation, with vendor participation.

Piloting of biological oxidation processes, using actual conversion process waters, is most strongly recommended in order to determine the actual residuals of ammonia, cyanide, thiocyanate, and other compounds. High surface area powdered activated carbon should be evaluated to establish the improvement in the lowering of residuals at different levels of sludge age and carbon content of sludge.

Granular carbon beds following biological effluent filtration should be piloted or tested in the laboratory to clearly establish the residuals of contaminants that may be reached by this method in comparison to the use of powdered activated carbon. Regeneration of powdered activated carbon in biological sludge by wet oxidation is another variation which merits investigation.

Since inorganic and organic removal is possible by reverse osmosis, it it recommended that removal of organic residuals, such as soluble oil, phenols, cyanides, and cyanates be investigated as well as removal of such inorganics as chlorides, boron, ammonia, and others.

The Parsons and COGAS Development Co. design concept of injecting stripped liquefaction wastewater into a high temperature gasifier which produces no phenolics, in order to destroy the organic impurities, should be tested in an integrated liquefaction pilot or demonstration plant. Investigation should include evaluation of possible problems from the inorganic components of the wastewater, such as catalyst plugging or poisoning and scaling in heat exchangers or pipes.

Long term experimentation, probably on the demonstration plant scale, is recommended for study of the effects on cooling tower operation of residual amounts of ammonia, cyanides, cyanates, sulfides,and various inorganic or organic compounds that may not be removed from the wastewater that is used for tower makeup. Supplying cooling tower vendors and specialists with actual samples from pilot plant operation would allow certain tests to be made in their laboratories. Certain parts of the problem are amenable to calculation such that recommendations on programs for prevention of corrosion and scaling could be made for the demonstration plant.

The possibility of establishing joint water treating programs between DOE and EPA at the DOE pilot plants should be investigated. Laboratory treatment studies, such as that being conducted by Carnegie-Mellon University under DOE sponsorship, might be extended. In the H-Coal demonstration plant, where the wastewater is to be treated and discharged without reuse, recycling schemes could be piloted. The water treating facilities at the rebuilt Conoco plant in Cresap, W. Va., although not clearly defined, may offer opportunities for experimentation.

Continuation of the EPA program at the University of North Carolina for treatment of conversion plant effluents is recommended, since the program is expected to yield useful information on the effects in treatment processes of specific substances found in conversion plant effluents.

Plans are needed for demonstration of the individual commercial water treating processes and the integrated schemes that use the processes. Treating results of virtually every step should be verified by actual testing on wastewaters from the processes involved. Although some of the steps may be sufficiently verifiable on wastewaters from pilot plants, the ultimate confirmation should come from integrated operation on a demonstration plant. The demonstration plant presumably would be large enough so that operations would be directly relatable to a commercial plant in every respect.

Demonstration of "zero discharge" water treating schemes employing recycle of treated water should be incorporated in demonstration plant designs. Specialists in cooling tower and boiler operations should be given subcontracts to participate in the design of these systems and also to monitor the systems during demonstration plant operation. In this way the best additives and conditions for control of scaling, algae, foaming, and corrosion could be established in the particular plant to be operated. Different plant locations, different raw water and coal compositions, and the different coal conversion processes, may present unique problems which could dictate additional equipment or different additives to control corrosion, scaling, algae, and foaming.

Study of efficiencies and costs for the following alternate commercial control technology for operation and actual corrosion process effluents is recommended:

- o Reverse osmosis on sludges with evaporation of reject stream. Treated water to cooling tower

- o Reverse osmosis on sludges with evaporation of reject stream. Treated water to demineralization for high pressure steam boiler feed water use

- Study of stage-wise water condensation to concentrate inorganic materials with a minimum of organic materials

- Establish best side-stream treatment system for cooling tower

- Establish best system for biological oxidation. The many variations offered include trickling filters, rotating biological disc contactors, fluidized sand beds, and High Purity Oxygen Activated Sludge. Licensors for the above have been documented and could be given samples and subcontracts sufficient to establish efficiency, capital cost, and operating costs on a firm-bid basis

- Anaerobic digestion as a first stage in biological oxidation, followed by an aerobic second stage. One licensor of this technology has been identified and others are available

- Powdered activated carbon addition to activated sludge systems. High surface area carbon and long sludge age, without regeneration, is one alternate. Regeneration by wet air oxidation and higher PAC rates is another. This system definitely enhances nitrogen compound removal as well as BOD and COD removal vs. conventional activated sludge with no carbon addition

- Thermal oxidation of wastewater (Zimpro) and catalytic oxidation of wastewater should both be tried. The final cleanup step following these must be established (probably biological oxidation at second stage conditions, preferably with powdered active carbon addition). High Purity Oxygen Activated Sludge (UNOX) is another candidate for the cleanup stage

- Establish the role, if any, of chemical oxidants such as ozone, chlorine, or hydrogen peroxide

Investigation into the efficiencies and costs of the following processes operating on actual conversion process effluents is recommended:

- Biological oxidation in fluidized bed (Ecolotrol, Inc.)
- Oil fluidized evaporation (Dehydrotech Corp.)
- Heavy metal removal by SULFEX process (Permutit)
- Ultraviolet irradiation with ozone
- Lime sludge recovery (Dorr-Oliver and others)
- Catalytic sludge precipitation (Permutit)
- Freezing (Fluor)
- Thermal incineration in reducing atmosphere and recycle to process of purge liquors from redox systems such as Stretford and Takahax (Nittetu Chemical Engineering, Ltd.)
- Coalescence of emulsified oil-water mixtures in solid beds. Substitute for API separator
- Lindman precipitator (Precipitator, Inc.). Uses SO_2, lime and iron to remove suspended solids, BOD, oil and grease
- Super bacteria strains for greater cleanup in biological oxidation (Polybac Corp.)

- Mercury removal processes (Japanese "re-elixirization," FMC, Georgia-Pacific)

FOR STUDIES OF GASEOUS EMISSION CONTROL

Sulfur Recovery vs. Sulfur Discard

Studies are recommended to develop data for economic decisions on:

- Reduce sulfur by physical coal cleaning and recover the rest of the sulfur in the coal conversion plant
- Reduce sulfur by physical coal cleaning, recover part in the conversion plant and discard the rest as FGD sludge
- Reduce sulfur by chemical (Meyers process) coal cleaning, recover or throw away the sulfur and recover or throw away the remaining coal sulfur in the conversion plant
- Recover all sulfur in the conversion plant
- Recover part of the sulfur in the conversion plant and throw away the rest

Particulates

Further study and evaluation are needed on the quantities and compositions, including particulates, of gases released from coal feeding devices such as lock hoppers.

Ash Quench

Much data and information are available on gases evolved from quenching in Lurgi Dry Ash gasification, but little real data have been collected on the gases evolved from other processes. Data are needed on quantity and composition of the ash quench streams from the various conversion processes and the variations for any one process as the quench water composition varies.

Acid Gas Removal

Removal of hydrogen sulfide and carbon dioxide from the process gas stream is not usually considered to be an emissions problem, but the composition of the hydrogen sulfide stream affects the performance of the sulfur recovery system, and thus affects the treatment that must be applied so that the final vent gas stream will meet environmental standards. In a like manner, the composition of the carbon dioxide stream affects the performance of downstream process steps and thus affects the vent gas treatment step.

Recommended studies in several of the acid gas removal processes are:

- Selexol: Study effect of feed acid gas composition on removal efficiency at various operating temperatures and pressures
- Rectisol: Determine the solvent retention of heavy hydrocarbons and the composition and quantity of the fugitive carryover from the process at high pressures

- Monoethanolamine (MEA): Study the effects of operating conditions on the formation of non-regenerable compounds, on excessive solvent losses, on corrosion and on foaming

- Diisopropanolamine (DIPA): Study the effect of operating pressure on hydrogen sulfide removal efficiency

- Diglycolamine (DGA): Study the effects of operating conditions on the formation of non-regenerable compounds and the effect of feed acid gas composition on removal efficiency at various operating temperatures and pressures

- Diethanolamine (DEA): Study means of removal of the fine particles that cause foaming, as removal efficiency vs. operability vs. cost. Data are needed on utilities requirements vs. operating temperature and pressure

- Fluor Solvent: Determine utility requirements and study effect of feed gas composition on removal efficiency at various operating temperatures and pressures

- Sulfinol: Determine solubility of hydrocarbons in the sulfinol solvent and study process economics vs. operating parameters

- Estasolvan: Study the effect of operating pressure on acid gas removal efficiency. Study methods of treatment for the blowdown stream

- Benfield: Study the process when it is operated selectively for the hydrogen sulfide content in the carbon dioxide stream. Determine the extent of COS hydrolysis vs. the requirements for Stretford process feed

- Amisol: Determine utility requirements and study effect of feed gas composition on removal efficiency at various operating temperatures and pressures

Sulfur Recovery and Tail Gas Cleanup

Further study of the sulfur recovery and tail gas cleanup process should include characterization of inlet and outlet gas streams, vent streams, byproducts, sulfur removal efficiency vs. operating parameters,and reactant degradation. Suggested areas of investigation in examples of processes include:

- Claus process: Determine extent of removal of HCN and ammonia from the feed gas stream and fate of CO and hydrocarbons in the feed gas. Study the effect on sulfur conversion of the presence of oxidizable compounds in the feed gas. Determine the economics of the process for operation on various feed gas compositions, with particular emphasis on the effects of variations in hydrogen sulfide concentration

- Stretford and Beavon processes: Determine conversion of organic sulfur compounds in the presence of high concentrations of carbon dioxide. Characterize oxidizer vent gas stream and solvent blowdown stream. Determine degree of removal of mercaptans and ammonia

Sulfur Dioxide Control

Data are needed on the effects on sulfur dioxide removal and on the process economics of changes in feed gas composition. In the Wellman-Lord process data on HCN and ammonia removal efficiency are needed. Data on composition and means of disposal of the process products are needed for all processes.

Hydrocarbon Control

Study of reduction or elimination of hydrocarbons in waste gas streams by incineration, absorption or adsorption is needed. Incineration in a utility boiler may affect boiler design and economics. Absorption into a solvent has been proposed but more information on the process types, efficiencies and economics is needed as these are affected by stream composition and quantity. The same information is needed for carbon adsorption.

Nitrogen Oxides Control

Studies of means of applying combustion modification, fluidized bed combustion, hydrodenitrogenation of liquid fuels and flue gas cleaning are needed to determine efficiency of removal and costs in coal conversion applications, with particular emphasis on incinerator/boilers. Investigations might be started in pilot plants and extended later to demonstration units.

FOR SOLID WASTES DISPOSAL AND MANAGEMENT

Coal Dust Suppression vs. Elimination

Determination of the size distribution of coal as it leaves the cleaning plant to determine the dust content, dust losses during transit to the coal conversion plant, effectiveness and costs of dust suppression in transit and effectiveness and costs for dust suppression at the coal conversion plant, as these are affected by coal type and source, should be carried out to establish the base case for suppression. The effectiveness and costs of dust agglomeration should next be determined. Then the costs and efficiency of dust control by dust agglomeration as a means of eliminating coal dust can be contrasted with the present practice of dust suppression.

Solids Waste Disposal

Studies on application of impervious liners to areas for disposal of ash and other inorganic coal conversion solids are recommended. Costs for application of the liners can then be determined as part of a recommended study on the economics of solid waste disposal on level ground, in strip mines or in valleys. Long term investigations on the effect of ash leachates on the proposed liners is recommended.

The acceptability of chemical stabilization of solid wastes should be determined in view of present and possible future environmental standards. Experimental work on ash/slag and cost development for the available stabilization processes is needed to establish process viability.

Systems for leachate collection and return to the conversion plant for reuse should be developed. Finally, the costs of liner application can be contrasted with chemical stabilization in view of environmental standards.

SECTION 5

CURRENT TECHNOLOGY BACKGROUND

DEVELOPMENT OF THE DATA BASE

Information concerning quantity and composition of the various emission, effluent, and waste streams from coal conversion processes was gathered by literature searches and by contacts with conversion process operators in order to define in detail the problems that must be solved so that the conversion processes can operate successfully without unacceptable deterioration of the environment. As they became available, data and information developed in the EPA program "Level 1 Environmental Assessment" were added to further define the problems of control.

In a parallel effort, Federal, State, regional, and international environmental standards were gathered to serve as a starting point for objectives for application and evaluation of control technology. As they became available, data and information developed in the EPA program "Multimedia Environmental Goals" were compared to the present and proposed environmental standards to aid in defining possible future control technology needs.

In a second parallel effort, information and data were gathered on available and developing control technology to define effectiveness and costs of controls that may be applied to conversion process streams so that the final streams leaving the process site meet environmental standards.

Information Procurement, Storage and Retrieval

Interrogation of key word systems concerning the various aspects of coal conversion yielded abstract lists from the National Technical Information Service (NTIS), the EPA library, Chemical Abstracts and the Franklin Institute. These were supplemented by searches through indexes of publications of the American Chemical Society and the American Institute of Chemical Engineers. Pertinent articles and reports ordered from these abstracts and indexes formed the nucleus of the project literature file. The annual, semi-annual, and weekly abstract publications of NTIS and EPA and various technical journals were monitored in a continuing effort to expand and update the literature file.

A computerized system was established for information storage and retrieval to provide a simple and efficient means of access to the literature according to an established keyword listing. The system operates with author, title, and category and does not include abstracts.

Subjects Monitored

The engineering personnel assigned to the project were divided into three groups with emphasis on:

Water Pollution Control
Air Pollution Control
Solids Disposal

Information on coal gasification and liquefaction processes was gathered in the initial stages of the project by all groups with particular reference to their assigned areas. Coordination was maintained by exchanging reports daily and in periodic meetings. Following the familiarization period, the three technical groups evaluated the coal conversion processes for high Btu gasification, low Btu gasification and liquefaction in terms of status (conceptual, laboratory, bench scale, process demonstration unit, commercial), chances for further development if less than commercial, and availability of data at present or forthcoming. From this evaluation the following list of processes was selected for particular attention:

High Btu Gasification:
- CO_2 Acceptor
- Bi-Gas
- Battelle Agglomerating Ash
- COGAS
- Synthane
- Lurgi Dry Ash
- Lurgi Slagging Ash
- Hydrane
- Hygas Steam/Oxygen
- Hygas Steam/Iron
- Texaco

Low Btu Gasification:
- Winkler
- Koppers-Totzek
- Westinghouse
- Foster Wheeler
- Combustion Engineering
- Riley-Morgan
- Wellman-Galusha
- U-Gas
- Babcock & Wilcox
- AI Molten Salt
- Morgantown*
- Bituminous Coal Research
- Woodall-Duckham

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*Morgantown Energy Research Center

Liquefaction: COED
Clean Coke
SRC I and II
H-Coal
Donor Solvent
Synthoil

Published analyses of wastewater, ash, char, coal, and other solids streams, and gas streams released to atmosphere, were collected by each group for their area of interest where data were available for the various process pilot or bench-scale units. Commercial scale conceptual designs prepared for many of the leading processes were collected for study. Quality of the conceptual designs varied depending upon contractor and degree of detail available from the licensor. Only the latest designs appeared to fully recognize the importance of pollution control.

The water group documented water treating system design and cost where available. The air and solid groups assembled similar information.

Personal Contacts, Trips and Meetings

The literature survey was supplemented by communication with technical paper authors, EPA and Department of Energy (DOE) project managers, operating personnel, control process vendors, and EPA and DOE contractors. A partial list of these contacts includes the following:

Carnegie-Mellon University
C. F. Braun & Company
Envirotech

Pittsburgh Consolidation Coal Company
Pittsburgh & Midway Coal Mining Company
Bituminous Coal Research
[illegible]hydro-Tech Corporation
Pittsburgh Energy Research Center (PERC)
Dravo Corporation
U. S. Steel
Texaco Development Corporation
Exxon Research & Engineering Corporation
Davy Powergas
Combustion Engineering
City Public Service, San Antonio
Morgantown Energy Research Center (MERC)
Foster Wheeler
Battelle
Institute of Gas Technology
EPA prime contractors: Catalytic, Radian, Hittman, Battelle, Hydrocarbon Research, Versar
Other EPA contractors in the Fuel Process Branch: North Carolina State University, University of North Carolina, Research Triangle Institute, Cameron Engineers, Illinois State Geological Survey, and Water Purification Associates
DOE project managers for SRC, Hygas, Clean Coke, Bi-Gas, CO_2 Acceptor, Donor Solvent
National Conference on Treatment and Disposal of Wastewater Residues
International Conference on Coal Gasification, Liquefaction and Conversion to Electricity
Pacific Chemical Engineering Conference
EPA Symposium on Environmental Aspects of Fuel Conversion Technology
Water Pollution Control Federation
Synthetic Pipeline Gas Symposium

American Institute of Chemical Engineers
National Coal Association
Purdue Waste Conference
Coal Waste Seminar

COAL GASIFICATION PROCESSES AND DATA GATHERING

A total of 24 coal gasification processes was chosen for investigation on the basis of their importance or stage of development. Of these, eleven processes are described as "high Btu" processes and thirteen as "low Btu" processes. Generally the low Btu processes are so described because air instead of oxygen is used; however, some of these do, or can, use oxygen, but the product gas is not methanated to Synthetic Natural Gas (SNG) for pipeline sale. All the high Btu processes produce SNG.

High Btu

- CO_2 Acceptor
- Bi-Gas
- Battelle
- COGAS
- Synthane
- Lurgi Dry Ash
- Lurgi Slagging Ash
- Hydrane
- HyGas Steam/Oxygen
- HyGas Steam/Iron
- Texaco

Low Btu

- Winkler
- Koppers-Totzek
- Westinghouse
- Foster Wheeler
- Combustion Engineering
- Riley-Morgan
- Wellman-Galusha
- U-Gas
- Babcock & Wilcox
- AI Molten Salt
- Morgantown
- Bituminous Coal Research (BCR)
- Woodall-Duckham

These processes may also be classified according to the design of the gasifier as:

Entrained Flow Gasifiers	(high-velocity, high carry-over of solids)
Fluidized Bed Gasifiers	(low velocity, such that a fluid bed is definable and solids carry-over is minimal)
Fixed-Bed Gasifiers	(large-particle coal feed, virtually no carry-over of solids)

Entrained Flow
Bi-Gas
Koppers-Totzek
Foster Wheeler
Combustion Engr.
Babcock & Wilcox
Texaco

Fluid Bed
CO_2 Acceptor
Battelle
COGAS
Synthane
Hydrane
HyGas (2)
Winkler
Westinghouse
U-Gas
BCR

Fixed Bed
Lurgi (2)
Riley-Morgan
Wellman-Galusha
Morgantown
Woodall-Duckham
AI Molten Salt

Entrained Flow Gasifiers

In this design, finely-divided coal is fed into a rapidly moving gas stream and remains suspended in cocurrent flow through the gasifier. Gasification is rapid because of high reaction temperatures and ash is removed as molten slag. The main advantages of this gasifier type are the ability to process all types of coal and the absence of tars and hydrocarbons higher than methane in the product gas (methane content is usually less than one percent by volume). The main disadvantages are the necessity to pulverize the coal feed and the possible carry-over of molten slag.

Bi-Gas--

The Bi-Gas process is funded jointly by DOE, the American Gas Association (AGA), Phillips Petroleum and Bituminous Coal Research (BCR). The 120 TPD (ton per day) pilot unit in Homer City, Pa. has been visited. Operation of the pilot unit has been very sporadic and no environmental data of consequence are available. Should it successfully operate, it will be extensively sampled by Penn Environmental Engineers (consultants hired by BCR) and by a team from Carnegie-Mellon University funded by DOE and AGA. This process contemplates the highest operating pressure of all the gasification processes at 80 to 100 atmospheres. The gas exit temperature is about 925°C and the bottom temperature is about 1,480 to 1,650°C. Predicted holding times are 2 seconds in the slagging ash zone and 6 to 8 seconds in the entrained bed section. Except for molten slag that is quenched and removed from the bottom section, all products, including char, are carried overhead.

Char recycle is planned, but has not been attempted so far in the pilot unit. The slagging ash section design is attributed to Babcock & Wilcox. The glassy ash removed is not expected to be leachable and discard to landfill is planned.

Koppers-Totzek--

This process is the only one of the entrained flow group that has been extensively commercialized. Twenty plants, many including more than one gasifier, have been placed in commercial operation in 14 countries since 1949. Operation is at substantially atmospheric pressure. Temperature of the gas leaving the gasifier is about 1.480°C. Temperature at the bottom (ash exit) is about 1,925°C. Useful data are available on liquid effluent streams.

A 150 TPD pilot unit, designed to operate at pressures up to 30 atmospheres, is scheduled for completion in late 1977. Sited in

Saarland, West Germany, the pilot plant will be funded and operated cooperatively by Shell Internationale Maatschippij, N. V. and Krupp-Koppers GmbH.

Texaco--

Another major effort on high pressure entrained gasifiers is being carried out by the Texaco Development Corporation in a 15 TPD unit at their Montebello, California research laboratory. Efforts to obtain environmental data have been unsuccessful, since this work is privately funded. It is known that Texaco's primary experience has been with heavy oil gasification and efforts have only recently turned to coal. The gasifier is reported to operate at pressures above 27 atmospheres and with reaction temperatures of about 1.650°C. Product gases leave a quench section at the bottom of the gasifier at 200 to 260°C. Texaco gasifier tests on residue from the Solvent Refined Coal (SRC) process have been reported.

A Texaco gasifier was reported to have been under consideration for COALCON, but COALCON itself, supposedly funded by DOE, has been cancelled. A 100 TPD pilot plant is being built in Germany by Ruhrkohle and Ruhrchemie. DOE has awarded a contract to W. R. Grace for development of a conceptual design to supply gas for ammonia synthesis. The Tennessee Valley Authority plans to build a similar, but smaller, combined plant. Other private organizations are interested in the process to generate gas for methanol synthesis and for power generation.

Foster Wheeler--

The Foster Wheeler gasifier is planned for operation on essentially the same principle as the Bi-Gas process, except that air is used instead of oxygen and the pressure is lower. Estimated pressure is 24 to 34 atmospheres. The gas exit temperature is about 980 to 1,150°C while the bottom temperature is about 1,370 to

1,540°C. A 480 TPD pilot plant is planned in association with DOE, Pittsburgh & Midway Coal and Northern States Power for completion in 1980, or possibly later. DOE and the University of Minnesota have awarded a contract for a gasifier to generate building heating steam. No environmental data have been reported for the process.

Combustion Engineering--
The Combustion Engineering process is planned for operation at atmospheric presure with a gas exit temperature of about 870°C and a bottom temperature of about 1,650°C. A pilot plant with EPRI and DOE involvement is scheduled for completion in mid 1978. No environmental data have been reported for the process.

Babcock & Wilcox--
Babcock & Wilcox gasifier designs resemble those of Koppers-Totzek and Texaco. It is reported that pressure experimentation will be carried out in the range of about 3 to 20 atmospheres. Reported temperatures are 980°C at the gas exit and 1,870°C at the bottom. Process design data have been published for a 480 TPD pilot plant that was scheduled for completion late in 1979 at the Seward Station of the Penn Electric Company. Apparently, funding was not forthcoming and the project was cancelled. No useful environmental data for the process have been reported.

Fluid Bed Gasifiers

The principal advantage claimed for fluid bed gasifiers is their isothermal operation. Oxygen consumption is reduced and, in the case of high pressure operation, the gasifier exit gas contains a higher portion of methane. The higher methane content increases the heating value as a fuel gas or reduces the extent of methanation facilities required.

CO_2 Acceptor--
The CO_2 Acceptor process, developed by Consolidation Coal Co. (CONOCO), is designed to process lignites and sub-bituminous coals. Bituminous coals cannot be used due to process limitations. An unusual feature is the use of recirculating dolomite to supply the heat for gasification reactions. Dolomite is regenerated in a separate vessel, where char is burned with air and the CO_2 absorbed in the gasifier is desorbed. The gasifier product gas contains substantial methane but no higher hydrocarbons or tars. Water analyses are available and the Carnegie-Mellon University team is committed to further water analytical effort for DOE and AGA. The 40 TPD pilot plant is no longer in operation. Pressure was about 10 atmospheres. The exit gas temperature was about 815°C and the regenerator temperature was about 1,010°C.

Battelle Agglomerating Ash--
Battelle/Union Carbide have developed a gasifier operating near slagging conditions that produces an agglomerated ash. A 25 TPD pilot plant began operation in mid 1977. Two fluidized beds are used. The first gasifies coal with steam alone at about 870 to 980°C, while in the second unreacted coal, char and recycled agglomerated ash are partially burned with air at about 1,090 to 1,150°C. Operating pressure is about 7 atmospheres. Environmental data are expected to be available during 1978.

COGAS--
Until recently, COGAS was a privately-developed process. A DOE contract awarded to the Illinois Coal Gasification Group now provides funds for preparation of a demonstration plant design to be located in Perry County, Ill. to produce 18 MMSCFD* of SNG

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*Million standard cubic feet per day.

plus 2,400 barrels per day of synthetic crude oil from 2,200 TPD coal. The design will be evaluated against the Lurgi Slagging Ash process to decide which process will receive matching DOE funding for construction and operation of a demonstration plant. COGAS coal gasification experimental data were developed in a 50 TPD pilot plant operating on char in Leatherhead, England. Very little process data are available, and there are virtually no effluent data. The quality of available data is good. If a COGAS demonstration plant is built, data will then be available. Some conceptual design data will be available when the Environmental Analysis Report is published, the date of which is not now known, but which is not expected before mid 1978.

Synthane--
Pittsburgh Energy Research Center (DOE/PERC) have developed the Synthane process, currently operating in a 75 TPD pilot plant at Bruceton, Pa. Pullman Kellogg visited the plant in June, 1977. There are some process difficulties with char let-down valve erosion and clinkering and build-up of fines. The process operates at about 68 atmospheres pressure with a top temperature of about 425 to 760°C and a bottom temperature of about 925 to 980°C. The goal is production of a high methane gas. The process uses steam and oxygen. In the original design a relatively large amount of phenols, oils, and tars (p/o/t) was produced. P/o/t production has been sharply reduced by injecting feed coal into the fluid bed rather than on top of it. Caking coal must be pretreated. Residual char is a problem, but it is intended that this will be used for power production and fuel in a commercial plant. The Carnegie-Mellon University team is monitoring the process for DOE and AGA, and PERC has published much analytical data from a smaller unit, including some water treating data.

Hydrane--
DOE/PERC is developing the Hydrane process in a small PDU (process demonstration unit). In this process coal devolatilization and pretreatment are to be accomplished in a free-fall dilute phase section of the gasifier. Char residue is used for hydrogen production. The methane content of the gasifier product gas is reported to be 35 to 55 percent, versus 10 to 15 percent for Synthane. Process operating pressure is about 68 atmospheres.

Temperature at the top of the reactor is about 540 to 820°C and about 980°C at the bottom. Little has been published on effluents from the process, but phenols, oils and tars are expected to be present.

HyGas--
Two HyGas processes were developed by the Institute of Gas Technology (IGT) in Chicago: steam/oxygen and steam/iron. The steam/oxygen process is further advanced and considerable effluent data from the 75 TPD pilot plant have been published. The Carnegie-Mellon University team is sampling and analyzing the effluents for DOE and AGA. Some water treatment data are near publication.

The steam/oxygen process is multi-stage with temperatures increasing from the top zone at about 315°C to the bottom zone at about 1,040°C. Operating pressure is about 82 atmospheres. Coal is fed as a slurry in recycled aromatic tarry hydrocarbons. Due to the relatively low temperatures of operation, the process produces phenols from all coals and oils when processing non-agglomerating coals. No tars are produced.

Other gasification methods tried were electrothermal char gasification (abandoned) and steam/iron. The latter is in the design stages for a 50 TPD pilot plant. This plant will also be monitored by the Carnegie-Mellon University team for DOE and AGA, but no data are yet available. C. F. Braun has completed a conceptual design for a commercial steam/iron plant, but evaluation shows substantially greater product gas cost than for steam/oxygen.

Winkler--
The Winkler process is fully commercial, with 16 plants, most containing more than one gasifier, having been built in Europe and Asia since the process was developed in the 1920's. Of these, four plants are still in operation. No reliable information on emissions is available. Davy Powergas are currently conducting an effluent characterization study on the operating commercial plants. The results of the study should be available in 1978. Davy Powergas have stated that the process does not produce phenols, oils,and tars. The process operates at substantially atmospheric pressure with a gas outlet temperature of about 705°C and a bottom temperature of about 815 to 980°C. Residence time in the gasifier is about 2 hours. An alternate design for operation at about 10 atmospheres is being developed.

Westinghouse--
This process was developed primarily to produce low Btu gas for power generation. A 12 TPD pilot plant is operating at Waltz Mill, Pa. A 120 TPD pilot plant is planned. A 1,200 TPD plant for Dresser Station in Terra Haute, Ind. is planned, but no completion dates have been announced. The process uses two beds, recirculating limestone or dolomite to remove sulfur and volatiles, and an agglomerating ash char gasifier. Coal is pretreated by a slipstream of fuel gas. Operating pressure is about 12 atmospheres. The desulfurizer operates at about 870°C and the gasifier at about 1090°C. Effluent data are scanty.

U-Gas--
U-Gas, developed by IGT, has been operated in a PDU mainly on coke and char feeds to demonstrate operability of the agglomerating ash process. The PDU was modified for coal feed and started in operation late in 1977. Developmental emphasis will be on the process, not on the effluents, until later in the program. Operating pressure may be varied from 1 to 27 atmospheres. Coal is pretreated, if necessary, at pressure and at a temperature of about 370 to 425°C. Gasification in the single bed is at about 1,040°C with 45 to 60 minutes residence time. The design features selective removal of high-ash material, a high carbon concentration in the bed, reinjection and combustion of elutriated fines and better than 95 percent carbon conversion. Little reliable emissions data are available.

W. R. Grace and Memphis Light are developing a conceptual design and feasibility study, under DOE contract, for production of 300 Btu per SCF pipeline gas from 2,800 TPD of coal.

BCR (Bituminous Coal Research)--
This 3-stage multiple fluid bed process has progressed into a PDU that has been operable since 1976. It is listed as a "major low Btu project" in DOE's coal conversion and utilization program. No reliable emissions data have been found. Operating conditions: pressures up to 16 atmospheres, first bed temperature 315 to 650°C, second bed temperature 925 to 1,090°C and third bed about 1,150°C.

Fixed Bed Gasifiers

Fixed bed gasification processes charge coal in lump form on a batch basis. During the reaction steam and either oxygen or air are charged to the bottom and gases move upward countercurrent to the slowly descending solids. This creates distinct temperature zones, from coal entry at the top of the bed to the ash forming zone at the bottom grate. Bottom temperature is usually kept below the ash fusion temperature of the particular coal, with the exception of the Lurgi Slagging Ash gasifier, while gas exit temperature may vary between 315 and 595°C, depending on coal characteristics and design variations.

Lurgi Dry Ash--

The Lurgi Dry Ash process best represents the commercial fixed bed gasifier technology. It is the only commercial process operating at high pressure. Eighteen plants have been built, most with multiple gasifiers, in nine countries. The gasifiers in Westfield, Scotland operate now only on a test basis. Operating pressure is about 24 to 31 atmospheres. Temperature at the top of the bed is about 370 to 540°C, about 650 to 815°C in the middle of the bed and about 980 to 1,370°C at the bottom. Because coal is fed to the top of the reactor and is gradually heated through stages of devolatilization, gasification and combustion, the gas leaving the reactor contains phenols, oils and tars.

Advantages claimed for the Lurgi Dry Ash process are high carbon conversion, low oxygen consumption, high throughput (due to high pressure), and relatively high methane make (7 to 12 percent by volume). Disadvantages cited are phenol, oil and tar production, top of bed clogging with fines, high steam consumption and difficulty with coals with low ash fusion temperature or those that cake or swell appreciably. A pretreatment for the latter (agglomeration) is available but consumption of oxygen and steam increase.

Virtually all the tentative commercial projects currently under consideration for the United States employ the Lurgi Dry Ash process. The following are the most widely publicized, but none have actually begun construction. Generally they are involved in the permit stage with public hearings on environmental issues, leasing agreement problems, etc. It is rumored that some (e.g., the Powder River Basin project) have been abandoned.

El Paso Natural Gas, Four Corners, N.M. Coal 28,250 TPD, SNG 288 MMSCFD (million standard cubic feet per day). Planning and feasibility studies continuing.

WESCO, Texas Eastern and Pacific Lighting, Four Corners, N.M., (4 plants). Each 25,625 TPD coal, 250 MMSCFD SNG. Environmental impact statement completed in early 1976. The latest report on WESCO (Oil and Gas Journal, March 6, 1978) states that the most recent difficulty is contract negotiations with the Navajo Indians.

Panhandle Eastern, Peabody Coal, Eastern Wyoming. Coal 27,700 TPD. SNG 270 MMSCFD. Environmental impact statement in preparation.

Natural Gas Pipeline of America, Dunn County, N. Dakota (4 plants). Each 30,000 TPD Coal, 250 MMSCFD SNG.

American Natural Resources Co., Peoples Gas Co. (North American Coal Gasification Corp.), Beulah-Hazen area, North Dakota. First phase, 137.5 MMSCFD gas, planned for 1982 completion. Second equal phase to follow. Environmental impact statement completed.

Northern Natural Gas, Cities Service, Powder River Basin, Montana (4 Plants). Each 30,000 TPD coal, 250 MMSCFD SNG.

Columbia Gas System (Illinois). 300 MMSCFD SNG.

Texas Eastern (Southern Illinois). 250 MMSCFD SNG.

The Lurgi process is proprietary and there are no emission data available for the gasifier itself. For the emissions from the total process, however, stream compositions are reported in conceptual designs, which in turn are based on real data. In addition, stream data were obtained for 4 American coals during runs on the commercial gasifier at Westfield, Scotland. An EPA sampling and analysis program on an operating unit in Yugoslavia is now in progress and should yield useful effluent and emission data.

Lurgi Slagging Ash--

This process, like the Lurgi Dry Ash, is proprietary and emissions data have not been published. The possibility of obtaining data from Conoco Coal Development Co., the proposer of the demonstration plant, appears to be small at present, but may improve later. DOE has awarded Conoco a contract to design the demonstration facility to produce 60 MMSCFD of pipeline gas from 3,800 TPD of coal. The process is more efficient than the Lurgi Dry Ash, requiring less steam and producing less aqueous effluents. The tentative conclusion may be drawn that because of the higher bottom bed temperature the gasifier exit gas will probably contain less phenols, oils and tars when operating on bituminous coals. (A gasifier operated by DOE at similar conditions at Grand Forks, S. D. with lignite feed shows high phenol, oil and tar in the gasifier product gas.)

Riley-Morgan, Wellman-Galusha--

These atmospheric pressure gasifiers have been fully commercial for many years. They may be operated to produce low Btu gas, using air, or medium Btu gas, using oxygen. Maximum bed temperatures range between 1,095 and 1,315°C. Depending on coal type, the exit gas temperatures range between 425 and 650°C and There are no dependable stream analyses available; however, phenols, oils,and tars have been reported as present in the emission streams. An EPA sampling and analysis program that is now in progress should yield useful effluent and emission information.

Morgantown (MERC/DOE)--
This 20 TPD day pilot plant at the Morgantown, Pa. Energy Research Center (MERC) is an extension of the Wellman-Galusha design embodying a stirred bed gasifier operating at pressures up to about 19 atmospheres. There are some published data on emissions from the process. There are conflicting reports of water treatment facilities planned for the pilot plant, but no treating data have been located.

Woodall-Duckham/Gas Integrale--
The Woodall-Duckham/Gas Integrale gasifier parallels the Wellman-Galusha, having been commercialized for over 30 years by the Italian firm Gas Integrale. Over 100 air-blown gasifiers have been successfully operated in Europe, South Africa, and Australia. Twenty-four oxygen-blown gasifiers have been operated in Europe. A typical air-blown unit has two stages. Approximately 200 TPD of coal are gasified to produce fuel gas and fuel oil (the fuel oil represents about 13% of the heating value). No analyses of tars or product water have been located. The latest reported activity concerns DOE negotiations with Erie Mining Co., Hoyt Lakes, Minn. to gasify 500 TPD of high sulfur agglomerating coals to produce fuel gas for iron ore furnaces.

Molten Salt--
The most nearly commercially proven gasifier in this category is the Otto-Rummel "Entrained Flow Slag Bath Gasifier" licensed out of West Germany. Gasifiers processing up to 250 TPD of coal were operated in the 1950's and 1960's. None are currently in operation, but a demonstration plant operating at about 25

atmospheres pressure was reported under construction in 1975. A clean gas is said to be produced, char is recycled and ash is disposed of by landfill. Bag filters for particulates collection and treatment facilities for process water are planned, but few details beyond this are available.

DOE is funding a 24 TPD molten salt gasifier developed by Atomics International. The plant is expected to be operational by early 1978 at Atomic International's laboratory in Santa Susanna, California. Since Pullman Kellogg has piloted a molten salt process of their own, some information has been exchanged with Atomics International.

Sulfur compounds in the coal react with the molten salt and the exit gas contains only about 5 ppm of total sulfur. Ash is removed as a moist filter cake and is washed free of soluble salts before it is carted to landfill. The melt bed operates at about 985°C. Exit gas temperature is about 925°C. Operating pressure may be varied from atmospheric to 19 atmospheres. No water is produced since no steam is fed to the process. No usable data on emissions have been published.

DEVELOPMENT OF GASIFICATION PROCESS EMISSION STREAM MODELS

The generalized block flow diagram of Figure 5-1 includes the major conversion process steps for production of high Btu gas. Since the concern is with emission, effluent, and waste (e/e/w) streams, recycle directly within the process is omitted and only those streams that would be expected to leave the process battery limits, or that must be treated before recycling, are indicated.

In the foregoing sections the efforts to collect data and information on the compositions and quantities of the e/e/w streams from the gasification processes were described and the lack of information for many processes and for streams within the processes was pointed out in the narrative. These results of the literature search and data gathering efforts become more apparent in the summarization shown in TABLE 5-1.

Inspection of the tabulation shows that for any one process there is insufficient information on the compositions and quantities of the e/e/w streams to allow specification of treatment methods for the streams. However, for most of the streams there are analyses, either real or derived from conceptual process designs, from one or more processes that could aid in defining the quantity and composition of the streams.

Gasification Process Categorization

The premise that conversion processes fed with the same coal and operating under the same or similar conditions will have the same or similar emissions was applied to the coal gasification processes. The groupings that resulted allowed application of emissions information among processes within each group in an attempt to close the information gaps.

Coal gasification processes were divided into processes in which little or no phenols, oils, and tars (p/o/t) are produced, and processes that produce p/o/t. The effect of the grouping on the availability of data within each group is shown in TABLE 5-2.

Those processes for which emissions data do not exist or are unavailable (Foster Wheeler, Texaco, COGAS, Westinghouse, U-Gas, BCR, Slagging Lurgi, Riley-Morgan, Woodall-Duckham) are omitted from the group lists. Although some doubt exists concerning future availability of usable information on some processes (Bi-Gas, Combustion Engineering, Battelle, Hydrane) these are not excluded from the categories. The AI Molten Salt process is

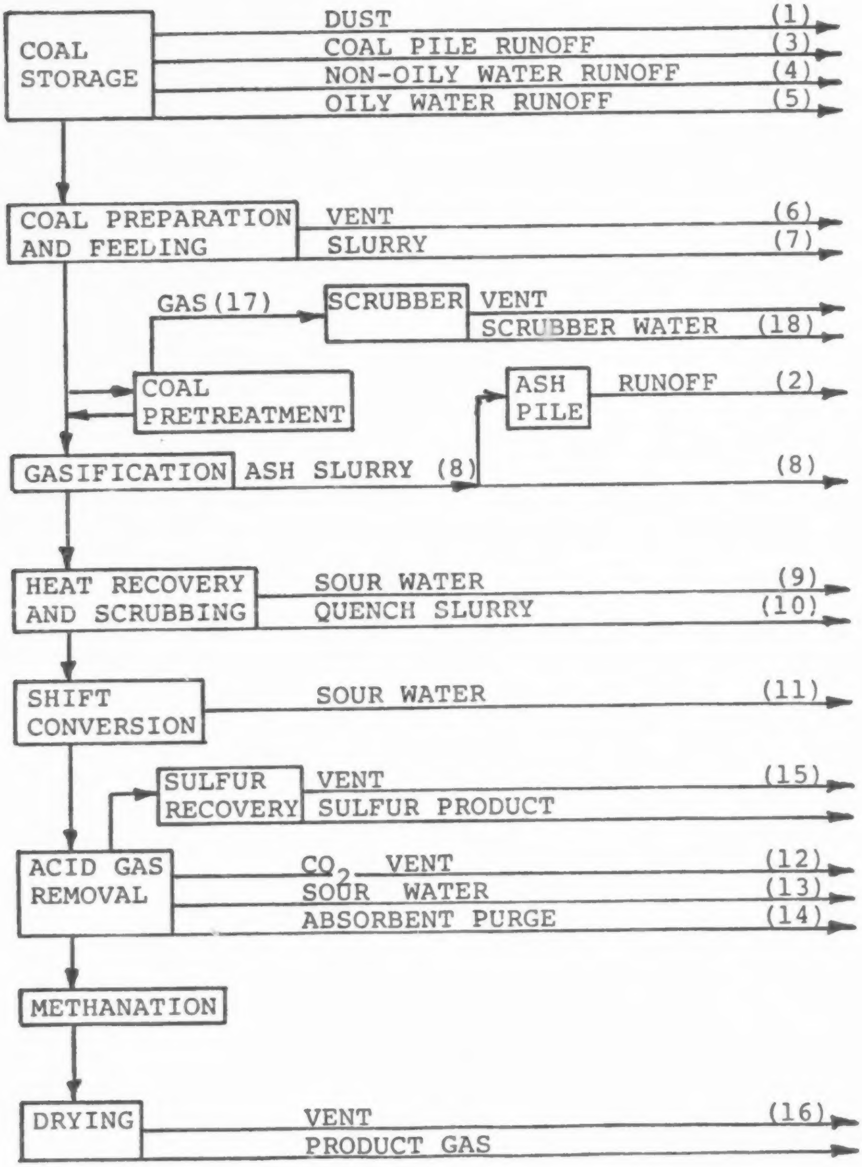
COAL STORAGE
DUST (1)
COAL PILE RUNOFF (3)
NON-OILY WATER RUNOFF (4)
OILY WATER RUNOFF (5)
COAL PREPARATION AND FEEDING
VENT (6)
SLURRY (7)
GAS (17)
SCRUBBER
VENT
SCRUBBER WATER (18)
COAL PRETREATMENT
ASH PILE
RUNOFF (2)
GASIFICATION
ASH SLURRY (8)
(8)
HEAT RECOVERY AND SCRUBBING
SOUR WATER (9)
QUENCH SLURRY (10)
SHIFT CONVERSION
SOUR WATER (11)
SULFUR RECOVERY
VENT (15)
SULFUR PRODUCT
ACID GAS REMOVAL
CO2 VENT (12)
SOUR WATER (13)
ABSORBENT PURGE (14)
METHANATION
DRYING
VENT (16)
PRODUCT GAS

Figure 5-1. Emission streams from coal gasification processes.

TABLE 5-1. AVAILABLE INFORMATION ON EFFLUENTS, EMISSIONS AND WASTES FROM COAL GASIFICATION PROCESSES

	Stream Analyses*																	
	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18
Bi-Gas								P		P	P				P	P		
Koppers-Totzek		A				A	A	P	A	A	A							
Foster Wheeler																		
Combustion Engr.								Q										
Babcock & Wilcox						A				P					A			
Texaco																		
CO_2 Acceptor							A	P	A	A	A					A		
Battelle								P		P	P				P	P		
COGAS																		
Synthane						A		P	A	A	A					A		
Hydrane								A										
HyGas (Steam/Ox.)							P	A	A	A	A	Q			Q	A		A
HyGas (Steam/Iron)						A	P	P	P	P	P				Q	A		
Winkler						A		P										
Westinghouse																		
U-Gas						Q		P			Q				Q			
BCR																		
Lurgi (Dry Ash)		A						A	A	A	A	Q			Q	Q		
Lurgi (Slagging)																		
Riley-Morgan								P			P							
Wellman-Galusha																		
Morgantown											P							
Woodall-Duckham																		
AI Molten Salt																		

*A = Analysis, either real or from conceptual design.

P = Partial analysis.

Q = Quantities

TABLE 5-2. CATEGORIZATION OF COAL GASIFICATION PROCESSES

	Stream Analyses (1)																	
	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18
Processes Producing No P/O/T																		
Bi-Gas								P		P	P				P	P		
Koppers-Totzek		A				A	A	A	A	A	A							
Combustion Engr.								Q										
Babcock & Wilcox						A				P					A			
CO_2 Acceptor							A	P	A	A	A					A		
Winkler						A		P										
U-Gas						Q		P			Q				Q			
CONSENSUS		A	(2)	(2)	(2)	A	A	P	A	A	A	(2)	(3)	(2)	A	A		
Processes Producing P/O/T																		
Battelle								P		P	P				P	P		
Synthane						A		P	A	A	A					A		
Hydrane								A										
HyGas (Steam/Ox.)							P	A	A	A	A	Q			Q	A		A
HyGas (Steam/Iron)							A	P	P	P	P				Q	A		
Lurgi (Dry Ash)		A						A	A	A	A	Q			Q	Q		
Riley-Morgan								P			P							
Wellman-Galusha											P							
CONSENSUS		A	(2)	(2)	(2)	A	P	A	A	A	A	Q	(3)	(2)	P	A		A

(1) A = Analysis, either real or from conceptual design.
P = Partial analysis.
Q = Quantity only.

(2) Data available from sources other than process descriptions.

(3) Stream 13 may be combined with Streams 9, 10 and 11.

excluded because its operating principles and e/e/w are completely different from those of the other processes.

Classifying gasification processes according to their production of p/o/t is useful because these components eventually appear in the waste water streams. Their presence requires the use of additional treatment units (for example, biological oxidation or phenol recovery) while their absence means significantly less intense water treatment will be needed. In addition, production of these contaminants generally reflects gasifier operating conditions, which in turn determines the form of solid waste produced (slag, agglomerates, or dry ash).

Phenols, oils, and tars may be formed during the gasification of coal. However, by increasing the gasifier temperature, the residence time or the average bed temperature (for example, by operating in the entrained flow mode or injecting the coal feed into the hot bottom part of the gasifier), production of phenols, oils and tars is reduced or eliminated.

It is noteworthy that the processes producing little or no p/o/t have either entrained flow or fluidized bed gasifiers that operate at temperatures of 1,035°C or higher and produce ash as a slag or as agglomerates. In contrast, the processes that produce p/o/t have either fixed bed or fluidized bed gasifiers operating at temperatures below about 1,000°C.

There are several exceptions to the generalization. The CO_2 Acceptor gasifier operates at less than 1,035°C but produces little p/o/t because the feed coal is injected into the bottom of the gasifier to yield a higher average bed temperature. Little is known at this time concerning the Battelle Agglomerating Ash process; however, sources indicate that no tars or oils are produced but that some heavy organics may be present in the gasifier exit gas.

An attempt was made in the early stages of the project to derive approximations of the quantities and compositions of the process streams in each of the two general process categories by averaging, or by exercise of best engineering judgement on, the available published data for the streams. It was conjectured that, used with caution, the deduced stream quantities and compositions would serve as a basis for evaluating means of application of available and developing control technology to the pollutants in the streams.

Analysis of the results of the attempt at deduction led to these conclusions:

- Data and information for coal storage, preparation, and feeding (streams 1 and 3 through 7) were, in most cases, fragmentary or derived from sources other than coal gasification.
- The consensus of the other streams from processes producing no phenols, oils, and tars (p/o/t) was derived almost entirely from Koppers-Totzek reports with a few additions from CO_2 Acceptor and Bi-Gas.
- The consensus of the other streams from processes producing p/o/t were derived principally from Lurgi Dry Ash, with some additions from HyGas and Synthane.
- Much information for all streams was based on conceptual engineering designs for commercial plants operating with particular gasification processes. Material balances in these process designs were

developed, in most cases, from laboratory, bench, and PDU data with attendant difficulties in scaling up the data and analyses to be representative of commercial operations. Combining stream analyses from different processes compounds the problem of deciding on stream compositions that can be considered to be typical for the processes within a category.

- The deduced stream compositions are at best only order of magnitude estimates or estimates of ranges of concentrations of the various constituents. Thus, if a design or evaluation of control technology happens to be especially sensitive to a precise concentration value, use of averages or ranges in the design or evaluation could result in errors.

Liquid Effluent Streams--
Consideration of the liquid effluent streams from processes producing p/o/t led to the general conclusions that data from Lurgi Dry Ash, possibly supplemented with Synthane and HyGas, designs may be used to characterize the streams and that control processes prescribed for treatment of the Lurgi, Synthane and HyGas streams will be applicable to similar streams from other conversion processes in the category. These conclusions are valid because Lurgi effluents contain most, if not all, of the contaminants to be found in effluent streams from any of the conversion processes in the category, and streams from other processes may differ from Lurgi streams in concentration or quantity but not widely in their components.

The same line of reasoning led to the conclusion that, for conversion processes producing no p/o/t, data from Koppers-Totzek operation, possibly supplemented by applicable CO_2 Acceptor or Bi-Gas data, could be used to characterize liquid effluent streams from this conversion process category. Accordingly, a block flow diagram was assembled to represent the liquid effluents from those processes producing p/o/t and a similar one was assembled to represent the liquid effluents from those processes producing no p/o/t. Both flow sheets show quantities and compositions of effluent streams. Assembly of these flow sheets and their uses are discussed in detail in Section 8 of this report.

Gaseous Emission Streams--
The characterization of the gaseous emission streams from the two categories of coal gasification processes was more complex for the following reasons:

- In order to specify the design characteristics of technology for control of tail gas emissions from the sulfur recovery section of the conversion process plant, the composition of the tail gas must be specified. It was EPA's intention that the tail gas composition would be developed by another contractor, but this was not done. Therefore, it was necessary for Pullman Kellogg to start at the acid gas purification step of the conversion process and determine its operation in high Btu gas production from low and high sulfur coal in order to ascertain the variations in the compositions of the feed gas to sulfur recovery and thus to determine the composition of sulfur recovery tail gas. Lurgi Dry Ash operation on low sulfur coal was taken as the base case. Alternate cases chosen were Lurgi operation on high sulfur coal and Koppers-Totzek operation, supplemented by CO_2 Acceptor and Bi-Gas data, on low and high sulfur coal.

- For the same reason as outlined above it was necessary for Pullman Kellogg to develop the variations in the composition of the vent gas stream from acid gas purification that may be anticipated with changes in conversion process and changes in coal feed in order to specify the technology required to control the emissions.

- Incineration of phenols, oils, and tars separated from the quench streams of those conversion processes producing p/o/t, and of the hydrocarbons in the acid gas purification vent stream, may be utilized to provide much of the heat required to raise steam for gasification. Accordingly, for these processes stack gases from operation of the incinerator as a utility boiler must be considered for specification of control technology, including final stack gas scrubbing. In those conversion processes producing no p/o/t, the incinerator requirement is greatly reduced and the heat is not available for process steam generation. Therefore, the steam must be raised by an offsite boiler fired with feed coal or product gas.

  Although development of control technology for boiler stack emissions is not considered to be part of this project, the assumption is made in the overall sulfur balances that sufficient H_2S is sent to the power boiler to react with the SO_2 in the boiler stack gases in a sulfur recovery scheme.

The flow sheets for the gaseous emissions base case and the alternate cases, together with detailed discussion of similarities and differences in the emission streams and the prescribed control technology, are contained in Section 9 of this report.

Solid Wastes--

Solids handling problems in coal conversion processes may be divided into two areas for attention, with the realization that these problems are, in the main, common to both gasification and liquefaction processes. The two areas of concern are the control of airborne dusts and the management of solid process wastes. Consideration of the paucity of data concerning quantities and compositions of such important streams as fugitive dusts and runoff waters, and of data concerning possible interactions between coal refuse and conversion process ash when these are mixed in a single disposal area, led to several conclusions:

- The fugitive dust problem may be solved by suppression of the dust or by collection and disposal of the dust. Lack of actual data, for example, on the amount of dust escaping from coal handling operations is no deterrent to investigation of means of controlling the dust.

- The hazardous and toxic properties of the runoff waters from coal storage piles, from piles of ash or ash plus coal refuse, and from dust scrubber slurry holding ponds can be dealt with by containment and recycling of the waters to prevent their entering the environment.

- Solid waste disposal may properly be considered as solid waste management. This change in viewpoint takes into account the means of adding solid wastes to the environment and the means of avoiding deleterious environmental effects.

Section 10 of this report develops the philosophies of control, containment and management of the solids in coal conversion processes into real solutions for the problems that minimize or eliminate environmental deterioration.

Base Case Flow Sheet and Material Balance

As pointed out in the foregoing discussion, the Lurgi Dry Ash process operating with a low sulfur coal feed, as described in the C. F. Braun 1976 report, (294, 295, 296), was chosen as the base case for determination of effluents, emissions, and wastes and the technology required for control of the pollutants. Supplemental information was derived from the Cameron Engineers' 1977 report (552) to close the material balance for the gas liquor separation, phenol extraction, and gas liquor stripping process steps. The block flow diagram of the process is shown in Figure 5-2. Discussion of the quantity and composition of the various process and effluent, emission and waste streams and their treatment, and comparison of alternate cases to this base case, will be undertaken later in the report.

The Lurgi plant design in the Braun report produces pipeline quality gas from a Western subbituminous coal. Coal from the mine is received in the coal handling section. The coal is crushed to less than 1-1/4 inches and is then screened with the greater than 3/16-inch fraction being sent directly to the gasification section while the less than 3/16-inch fraction is used as fuel in the main boiler.

Coarse coal in a size range of 1-1/4 inches x 3/16 inch is fed to the gasification section where it is gasified at a pressure of 450 psig with oxygen and high pressure steam. Ash from the moving-bed gasifiers is quenched with water. The total ash leaving the gasification section contains about 12.7 percent carbon.

The raw synthesis gas leaving the gasifiers is quenched with water and cooled in a waste heat boiler to produce low pressure steam at about 100 psig. Condensed tarry gas liquor is sent to the Phenosolvan unit.

Cooled synthesis gas is next sent to the shift conversion and cooling section where the H_2 to CO ratio is adjusted in preparation for methane synthesis. A cobalt-molybdenum catalyst is employed which is activated by H_2S and which also aids hydrogenation. Carbon monoxide is converted to hydrogen via the water gas shift reaction:

$$CO + H_2O = CO_2 + H_2$$

Also some COS is converted to H_2S via the reaction:

$$COS + H_2S = CO + H_2S$$

Small quantities of higher phenols and heavy tars are hydrogenated to lighter products. Some desulfurization of naphtha also occurs. It is necessary to by-pass about 45 percent of the gas around the shift converters to control the H_2/CO ratio. After shifting, the gas is cooled in waste heat boilers where low pressure steam at 125 to 150 psig is produced. The gas is further cooled in boiler feed water heaters and finally cooled in air- cooled and water-cooled exchangers. The condensed oily gas liquor is combined with that condensed in the gasification section and sent to the Phenosolvan unit.

In the Phenosolvan unit tars and tar oils as well as expansion gas are separated from the aqueous streams and routed to the main boiler. Clean and contaminated gas liquor streams from separation are sent to the extraction section. Phenols in the gas liquor from the gasification and gas cooling sections are extracted from the water with isopropyl ether as the solvent. The solvent is distilled from the phenols and returned to the

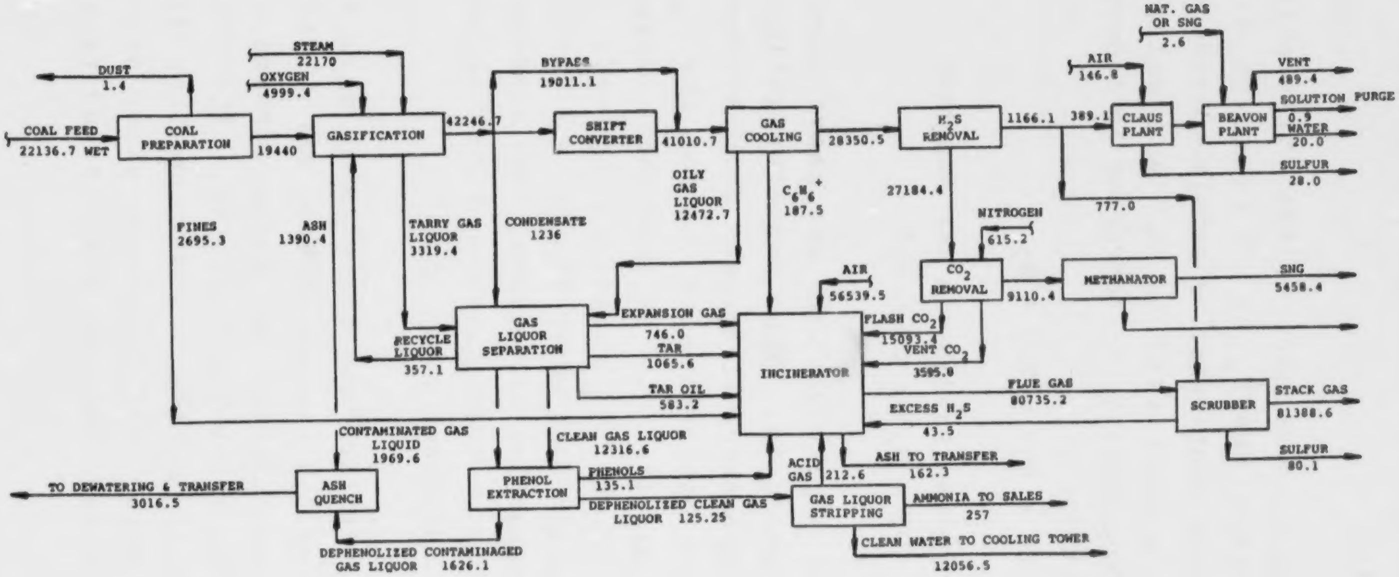
COAL FEED
22136.7 WET
DUST
1.4
COAL PREPARATION
19440
FINES
2695.3
STEAM
22170
OXYGEN
4999.4
GASIFICATION
ASH
1390.4
TARRY GAS LIQUOR
3319.4
42246.7
BYPASS
19011.1
SHIFT CONVERTER
CONDENSATE
1236
41010.7
GAS COOLING
OILY GAS LIQUOR
12472.7
$C_6H_6^+$
187.5
28350.5
H_2S REMOVAL
27184.4
1166.1
389.1
777.0
AIR
146.8
CLAUS PLANT
NAT. GAS OR SNG
2.6
BEAVON PLANT
VENT
489.4
SOLUTION PURGE
0.9
WATER
20.0
SULFUR
28.0
NITROGEN
615.2
CO_2 REMOVAL
9110.4
METHANATOR
SNG
5458.4
GAS LIQUOR SEPARATION
RECYCLE LIQUOR
357.1
EXPANSION GAS
746.0
TAR
1065.6
TAR OIL
583.2
AIR
56539.5
FLASH CO_2
15093.4
VENT CO_2
3595.0
INCINERATOR
FLUE GAS
80735.2
EXCESS H_2S
43.5
SCRUBBER
STACK GAS
81388.6
SULFUR
80.1
CONTAMINATED GAS LIQUID
1969.6
ASH QUENCH
TO DEWATERING & TRANSFER
3016.5
CLEAN GAS LIQUOR
12316.6
PHENOL EXTRACTION
PHENOLS
135.1
DEPHENOLIZED CONTAMINAGED GAS LIQUOR 1626.1
DEPHENOLIZED CLEAN GAS LIQUOR 125.25
ACID GAS
212.6
ASH TO TRANSFER
162.3
GAS LIQUOR STRIPPING
AMMONIA TO SALES
257
CLEAN WATER TO COOLING TOWER
12056.5

Figure 5-2.- Flow diagram for SNG production by Lurgi gasification of low sulfur coal.

extractor. The dephenolized water is then stripped with steam in the deacidifier column to remove CO_2 and H_2S. Acid gas from the top of the column is sent to the boiler and the bottoms flow to the ammonia stripper.

An aqueous ammonia stream is removed overhead from the ammonia stripper column and sent to storage. The bottom stream is relatively clean water which goes to further processing.

The cooled, shifted synthesis gas is heat treated to remove naphtha. This removal is effected by either chilling followed by vapor-liquid separation and reheat or by absorption of the naphtha (C_6H_6 and higher) in a lean oil followed by distillation.

The synthesis gas is next processed in two separate Selexol gas purification units. The first removes H_2S, CS_2, and a small percentage of the COS. The second removes CO_2 and the remaining COS. The Selexol units use organic solvent at medium to low temperatures.

The Selexol H_2S removal unit employs four absorption trains and two stripping trains to accomplish essentially complete removal of H_2S (1 ppm H_2S remains in the absorber overhead gas). Rich solvent from the absorbers flows through a series of flash drums to remove CO_2 and produce a concentrated H_2S stream. Flash gas is compressed and returned to the bottom of the absorbers after cooling. Rich solvent, now depleted in CO_2, is pumped to the strippers. H_2S is removed overhead and routed to the Claus plant for sulfur recovery. Lean solvent from the bottom of the strippers is pumped through heat exchangers and coolers before returning to the absorbers. Low pressure steam provides heat for reboiling the strippers while air and cooling water remove heat in the overhead condensers.

The Selexol CO_2 removal unit employs eight absorption trains and four stripping trains to accomplish about 95 percent removal of CO_2 and essentially complete removal of COS. Chilled lean solvent contacts feed gas countercurrently in the absorbers. The lean solvent temperature is optimized in the design for minimum power requirements for the system. Rich solvent flows through a series of hydraulic turbines (for power recovery) and flash drums. Most of the CO_2 (about 83.4 percent of that removed) along with all of the entering COS is released from the last flash drum. This stream contains substantial quanties of CO, H_2, and hydrocarbons (CH_4 and C_2H_6) and therefore must be treated before release to the atmosphere. Since the heating value is substantial, the recommended treatment is incineration in the main boiler.

Further in operation of the Selexol unit, semi-rich solvent is pumped to the CO_2 stripper where the remaining 16.6 percent of the CO_2 is removed by stripping with nitrogen gas. The presence of combustibles in this stream similarly dictates further treatment. Incineration in the main boiler is the method chosen in this study. Lean solvent from the bottom of the strippers is pumped back to the absorbers.

The CO_2 absorber overhead gas flows to sulfur guard beds which use zinc oxide to remove trace quantities of sulfur-bearing compounds. The purified synthesis gas proceeds to the methane synthesis section where the CO and H_2 are catalytically converted to methane via the following reactions:

$$CO + 3H_2 = CH_4 + H_2O$$

$$CO_2 + 4H_2 = CH_4 + 2H_2O$$

The conversion is accomplished in a three-stage, fixed-bed reaction system where the bulk of the methanation is done in the first two stages with recycle of effluent gas for temperature control. High pressure steam is produced in waste heat boilers following the methanators. The final heating value of the gas is reached in the third-stage menthanator. The gas at 970 Btu per SCF is produced at 285 psig. The wet product gas is compressed to 1017 psig, cooled to 100°F and sent to product gas drying.

Drying is accomplished in a conventional glycol dehydration unit. Water is absorbed from the gas stream and subsequently stripped out in the regenerator column. Overhead water vapor from the column contains some CH_4 and the stream must be incinerated. Substitute natural gas (SNG) at 100°F and about 1,000 to 1,015 psig is delivered to the pipeline.

COAL LIQUEFACTION PROCESSES AND DATA GATHERING

As with coal gasification, several coal liquefaction processes have been proposed and have been carried through some or all of the usual stages of development, as laboratory scale, process development unit (PDU), pilot plant, and demonstration plant. Only two, the Bergius system that was developed to produce fuel oil and gasoline and the Fischer-Tropsch system that was developed to produce a wide range of organic chemicals and fuels, have been operated commercially, and of these only the Fischer-Tropsch system is still in operation. Data and information on the emissions, effluents and wastes (e/e/w) from coal liquefaction processes other than the Fischer-Tropsch system must therefore be derived from small scale operations in the form of reports of operations, reports of sampling and analysis of e/e/w and from conceptual process designs that were prepared as economic and, in some cases, environmental impact studies.

Review of published literature on development of coal liquefaction processes revealed that most of the available reports were concerned principally with progress of process development and operating experience and that e/e/w were not treated in detail. Conceptual designs for commercial scale plants made use of as much e/e/w data as were available and calculated the remainder of the material balances. The problem of securing data and information on the e/e/w of some of the developing processes was further complicated by the fact that initial process development was private, the process information was considered to be proprietary and there were no publications of usable information.

Published information is available on six coal liquefaction processes that are judged to be of importance according to stage of development, schedules for construction and further development, potential for commercialization, probability of successful development and probable applicability of the process. In these processes operating conditions are chosen so that the residual solid produced can be minor, as an ash concentrate, or major, as low sulfur coal or metallurgical coke. All of these processes employ hydrogen in order to increase the yield of liquid products.

The six processes may be grouped according to their general type of operation as either pyrolysis/hydrocarbonization or solvent hydrogenation:

Pyrolysis/Hydrocarbonization
COED
Clean Coke

Solvent Hydrogenation

Solvent Refined Coal (SRC)
H-Coal
Donor Solvent
Synthoil

In general, coal liquefaction processes are more nearly alike than are coal gasification processes. For example, since all liquefaction processes produce hydrocarbon liquids, it is inevitable that there will be effluent streams containing phenols, oils, and tars (p/o/t) and that these streams will require effluent control systems similar to those applicable to the p/o/t-producing gasification processes.

Hydrogen for coal liquefaction is generated either by light hydrocarbon reforming or by gasification of residue or char. The general statement may be made that hydrogen production by similar methods yields similar effluents and requires similar control methods for that process step. All conceptual designs for commercial plants include gasification processes that produce no p/o/t, with operation at high temperatures and discharge of ash as a slag.

Coal Liquefaction Processes

COED--

The process uses a four-stage fluidized bed reactor usually operating at a pressure less than 1 atmosphere. The first stage temperature is about 175°C, the second is about 430°C, the third is about 540°C and the fourth is about 815°C. Oil, gas, and char are produced in the first three stages of pyrolytic liquefaction. The char is partially gasified in the fourth stage to produce the hydrogen for the process.

Early studies on the gasification of the COED char led to the development of the COGAS process which can operate with either the char feed or with coal feed. With char feed, COGAS may be considered to be combined with COED as the fourth stage of the process.

The process was developed by FMC and others. It was demonstrated in a 36 TPD pilot plant in Princeton, N. J. When testing was completed in 1975 the pilot plant was dismantled. Data from the operation were used in a conceptual design for a 25,000 TPD plant by Ralph M. Parsons Co. The COGAS process, developed to gasify the COED char, occupies the efforts of FMC and others. The project is now called the COGAS Development Company. The Illinois Coal Gasification Group is at present the main supporter of the project, together with DOE. Emissions data from COED are scanty and incomplete. As noted in the COGAS status description, although a DOE contract has been signed for development of a conceptual design, completion date for the design has not been announced.

Clean Coke--

Dry coal is pulverized to minus 20 mesh, then pyrolyzed in the carbonization section in a fluidized bed at about 705 to 760°C and at a pressure of about 7 to 10 atmospheres. In the hydrogenation section, hydrogen and a slurry of coal and oil are fed and processed at about 200 to 300 atmospheres. Char from the pyrolysis section is pelletized with process-derived heavy oil and is partly gasified to yield hydrogen for the pyrolysis and hydrogenation steps and the process product, metallurgical coke. Fuel gas and medium oil are other products.

U. S. Steel developed the process and operated a 1,000 pound per day process development unit on Illinois No. 6 coal. The project is now continuing with DOE participation. No emissions data have been published. A 100 TPD pilot plant has been announced as planned for Monroeville, Pa. but no completion date has been set. A conceptual design for a commercial plant is in progress.

Solvent Refined Coal (SRC)--
The SRC process is being developed by the Pittsburgh & Midway Coal Mining Company, a subsidiary of Gulf Oil, in Fort Lewis, Washington and is operating in a 50 TPD pilot plant with DOE participation. Dry coal, solvent, and hydrogen are fed to a dissolver operating at 68 to 136 atmospheres at temperatures from about 425 to 495°C. Final products from the dissolving section are solvent refined coal, fuel oil, and naphtha. In a commercial plant residue from the process would be gasified to produce process hydrogen and SNG.

Data from operation of the 50 TPD pilot plant at Fort Lewis, Washington have been reported in literature and progress reports. Unfortunately, much of the water treatment data is for the combined total of wastewater from the process plus once-through cooling water plus cooling tower and boiler blowdowns plus the plant runoff. The reported data demonstrate the treatability and biodegradability of the total, highly diluted SRC wastewaters, but do not allow consideration of control technology application to the individual streams. It is understood that plans are underway for more detailed analysis of the process emission streams. The process has been changed from the original, designated SRC I, to a variant designated as SRC II. It is understood that the process change will not significantly change the wastewater composition.

DOE is the principal participant in the Fort Lewis operations, with Pittsburgh & Midway operating the pilot plant and performing control analyses. Various DOE subcontractors have been, or are, conducting environmental and toxicity studies. Some usable emissions data have been collected and more information is being developed. SRC therefore appears to be one of the best sources of liquefaction information and samples for evaluation of control technology, provided DOE decides to let contracts to licensors and vendors of control technology and equipment.

H-Coal--

In Hydrocarbon Research's H-Coal process coal is hydrogenated directly in an ebullating bed of catalyst at a temperature of about 455°C and at a pressure of about 200 atmospheres. Liquid products are either low sulfur boiler fuel or synthetic crude oils, depending on the degree of hydrogenation. Gaseous product is SNG. Residue, in some cases supplemented by coal or light hydrocarbons, is gasified to produce process hydrogen.

The H-Coal process was developed by Hydrocarbon Research Inc. through bench-scale and PDU for production of synthetic crude oil. Recent studies have been in the conversion of coal into low sulfur fuel oil. A 600 TPD pilot plant is now under construction in Catlettsburg, Kentucky, scheduled for 1978 completion and for operation into 1980. DOE is funding about 75% of the work, with the remainder coming from industry and EPRI. Fluor is preparing a conceptual commercial design. Water analyses and useful treatment data have been obtained. Aware, Inc. was engaged to perform treating experiments and reported some of the results at the recent Purdue Waste Conference. The full text of the Aware report has been released to Pullman Kellogg by DOE.

Donor Solvent--

The Exxon Donor Solvent process requires coal ground to minus 30 mesh. This feed is slurried in recycle solvent that carries part of the hydrogen for the process and is mixed with more hydrogen in the liquefaction reactor that is operating at 99 to 172 atmospheres and at about 370 to 380°C. Actual liquefaction conditions are dependent upon the coal being processed and the degree of conversion desired. No catalyst is used in the liquefaction reactor.

Liquefaction product is separated by vacuum distillation into liquids, gas, and a residue consisting of unreacted coal and ash. The raw coal liquids are processed further as synthetic crude oil. Spent solvent is catalytically hydrogenated at (unspecified) temperatures and pressures for recycle to liquefaction. Gases from distillation and solvent hydrogenation are used for fuel or for hydrogen generation. The residual still bottoms are used to generate hydrogen and, in the course of the gasification reactions, produce more raw coal liquids.

Exxon is operating a one TPD PDU at Baytown, Texas. Proprietary designs are being developed for a 250 TPD pilot plant and for a commercial plant.

DOE is now sponsoring operation of the PDU. No emissions analyses and little specific process data were published by Exxon prior to DOE sponsorship. Although Exxon is reporting to DOE monthly, to date these reports have not become available through publication services. DOE is providing funding for half the construction and operation of the pilot plant, with Exxon and other industry sources providing the rest. Construction of the pilot plant in Baytown, Texas is scheduled to start in mid 1978.

Synthoil--
In a turbulent, cocurrent, upflow packed bed reactor, a slurry of coal and recycle oil is catalytically hydrogenated into gas and low sulfur fuel oil products. Operating temperature is about 450°C and pressure is about 136 to 271 atmospheres. Residue from liquefaction is used to produce the process hydrogen.

The process was developed originally by the U.S. Bureau of Mines and development is now continuing at the Pittsburgh Energy Research Center. A 10 TPD PDU is being built in Pittsburgh, scheduled for 1978 completion. Foster Wheeler Energy Corporation designed and is managing the construction of the PDU.

Several reports have been published by USBM and include process material balances, trace element distribution among the various process streams, evaluation of the environmental aspects of the process, and product analyses. Since most of the data concern the process and products, only a small portion of the data is usable for definition of control technology requirements. The trace element determinations were preliminary and exploratory in nature but do indicate that most of the trace elements appear in the residue and only very small amounts appear in the process effluent water.

DEVELOPMENT OF LIQUEFACTION EMISSION STREAM MODELS

The initial efforts in gathering data and information on the compositions and quantities of the emission/effluent/waste (e/e/w) streams from the liquefaction processes were not as successful as were those concerning gasification processes, principally because most of the processes are proprietary, or were until DOE participation began, and little usable process and e/e/w information was published by the process developers. A secondary reason for lack of information is the normal R&D emphasis on process development rather than on process e/e/w evaluation. Construction and operation of pilot plants, however, requires provision for treatment and disposal of e/e/w, with the added fact that such streams are large enough in the pilot plant operation to allow meaningful sampling, analysis, and quantity measurements to be taken. The future operations of the processes therefore should yield information usable in determining control technology needs designed for specific streams from specific processes. Until this information is available, however, the compositions of most of the e/e/w streams from coal liquefaction processes must be estimated from published data or by comparison with similar streams in other processes.

The generalized block flow diagram of Figure 5-3 includes the major process steps in coal liquefaction processes. Only those streams that would be expected to leave the process battery limits or that must be treated before recycling are indicated on the diagram. The results of the literature search and data gathering efforts are summarized in TABLE 5-3.

Emissions, Effluents and Wastes from Liquefaction Processes

The gaps in the e/e/w stream data that are shown in TABLE 5-3 obviate any attempt to derive approximations of quantities and compositions of effluents, emissions and wastes, in a manner similar to the early attempt in the gasification process analysis, and for much the same reasons.

Because of the similarities of the e/e/w from the various liquefaction processes, and because the differences in the streams between processes are principally in concentration and quantity and to a lesser degree in composition, the premise may be adopted that control technology applied to one liquefaction process should be generally applicable to the other processes. Accordingly, the Solvent Refined Coal (SRC II) process was chosen as representative of liquefaction processes and became the base case for the investigation of control of effluents, emissions, and wastes. Data from the Ralph M. Parsons Company conceptual design for SRC II (814) was supplemented by analyses and treating results from H-Coal to establish representative quantities and compositions of the streams.

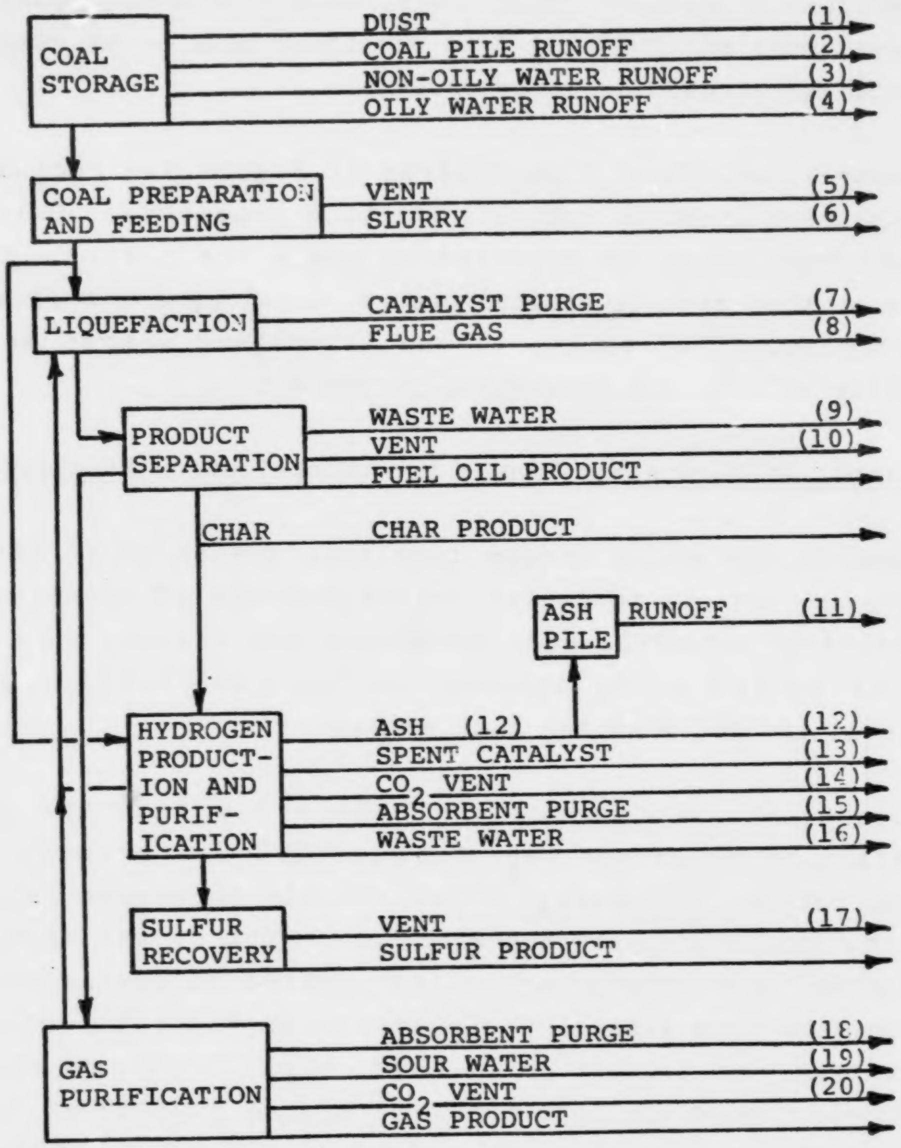
COAL STORAGE
DUST (1)
COAL PILE RUNOFF (2)
NON-OILY WATER RUNOFF (3)
OILY WATER RUNOFF (4)
COAL PREPARATION AND FEEDING
VENT (5)
SLURRY (6)
LIQUEFACTION
CATALYST PURGE (7)
FLUE GAS (8)
PRODUCT SEPARATION
WASTE WATER (9)
VENT (10)
FUEL OIL PRODUCT
CHAR
CHAR PRODUCT
ASH PILE
RUNOFF (11)
HYDROGEN PRODUCTION AND PURIFICATION
ASH (12)
(12)
SPENT CATALYST (13)
CO_2 VENT (14)
ABSORBENT PURGE (15)
WASTE WATER (16)
SULFUR RECOVERY
VENT (17)
SULFUR PRODUCT
GAS PURIFICATION
ABSORBENT PURGE (18)
SOUR WATER (19)
CO_2 VENT (20)
GAS PRODUCT

Figure 5-3. Emission streams from coal liquefaction processes.

TABLE 5-3. AVAILABLE INFORMATION ON EMISSIONS FROM COAL LIQUEFACTION PROCESSES

	Stream Analyses*																			
	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20
COED			P						P			P		Q		P	A		P	A
Clean Coke																				
SRC	Q								A			Q					A			A
H-Coal							Q		A							A			A	
Donor Solvent																				
Synthoil									Q			A								

* A = Analysis, either real or from conceptual design.
P = Partial analysis.
Q = Quantities only.

Base Case Flow Sheet and Material Balance

The block flow diagram of the SRC II liquefaction process is shown in Figure 5-4 with the weight flows of the various process and effluent, emission, and waste streams indicated. Discussion of the composition of the streams and their treatment, and comparison of alternate treatment means to the base treatment will be undertaken in Section 8 for liquid effluents, Section 9 for gaseous emissions and Section 10 for solid wastes.

The SRC II plant design in the Parsons report produces naphtha, fuel oil, pipeline quality gas and other gases equivalent to liquefied petroleum gases. Coal from the mine is received in the coal handling section and ground to about 800 micrometers. The ground coal is mixed with solvent, hydrogen is added and the mixture is heated and fed to the dissolver where, at 150 atmospheres pressure and at 455°C, the coal reacts exothermally and dissolves. The dissolver product is separated into a gas phase and a slurry phase. The gas phase is cooled, depressurized, and separated into a gas stream, which is fed to the acid gas removal section, and a liquid stream, which is further separated into a hydrocarbon stream and a sour water stream.

The slurry phase from the dissolver, containing dissolved coal, unreacted coal, char, and dissolved gases, is cooled, depressurized, mixed with recycled wash oil from the filtration unit, and fed to a fractionator for separation into light and heavy distillates and bottoms. Light ends are stripped from the distillates and hydrogenated to naphtha at about 50° API gravity. Some of the heavy distillate is blended with the light distillate and the filtrate from the filtration unit into fuel oil product. The rest of the heavy distillate is used as wash oil in the filtration unit. The fractionator bottoms stream, containing all of the solid residue from liquefaction, is cooled and divided, with part being recycled to the coal slurrying step as process solvent and the rest being filtered to separate the solids as filter cake from the liquid. The liquid is recycled to the coal slurrying step. The filter cake is dried and fed to the fuel gas gasifier.

The coal dissolver gas phase, after cooling, separation, and treatment to remove acid gases, is dried and cryogenically separated into a hydrogen-rich gas stream, a methane-rich gas stream and a vapor-liquid stream containing ethane and heavier hydrocarbons. Part of the hydrogen-rich stream is recycled to the coal disolver while the rest is treated in a methanator, to convert its carbon monoxide to methane and water, then dried and sent to naphtha hydrogenation. The methane-rich stream is methanated, compressed, cooled, dried, and adjusted in composition to the SNG heating value of 1,050 Btu per SCF (HHV)* by admixture of ethane and propane from the deethanizer in the LPG fractionation unit. In the LPG fractionator the vapor-liquid stream is separated into ethane for recycle and propane and butane as LPG products.

In a 2-stage, entrained flow, slagging gasifier, operating at 1,000°C, coal is converted to methane and synthesis gas. The slag residue is quenched, partially separated from the quench water in hydrocyclones, and then sent to disposal as a slurry or as wet solids. The excess quench water is cooled and recycled. Product gas is cooled, particulates are separated in cyclones and dust filters and returned to the gasifier, then the clean gas reacts with steam over a sulfur-resistant catalyst to hydrogen and

*Standard cubic foot (higher heating value)

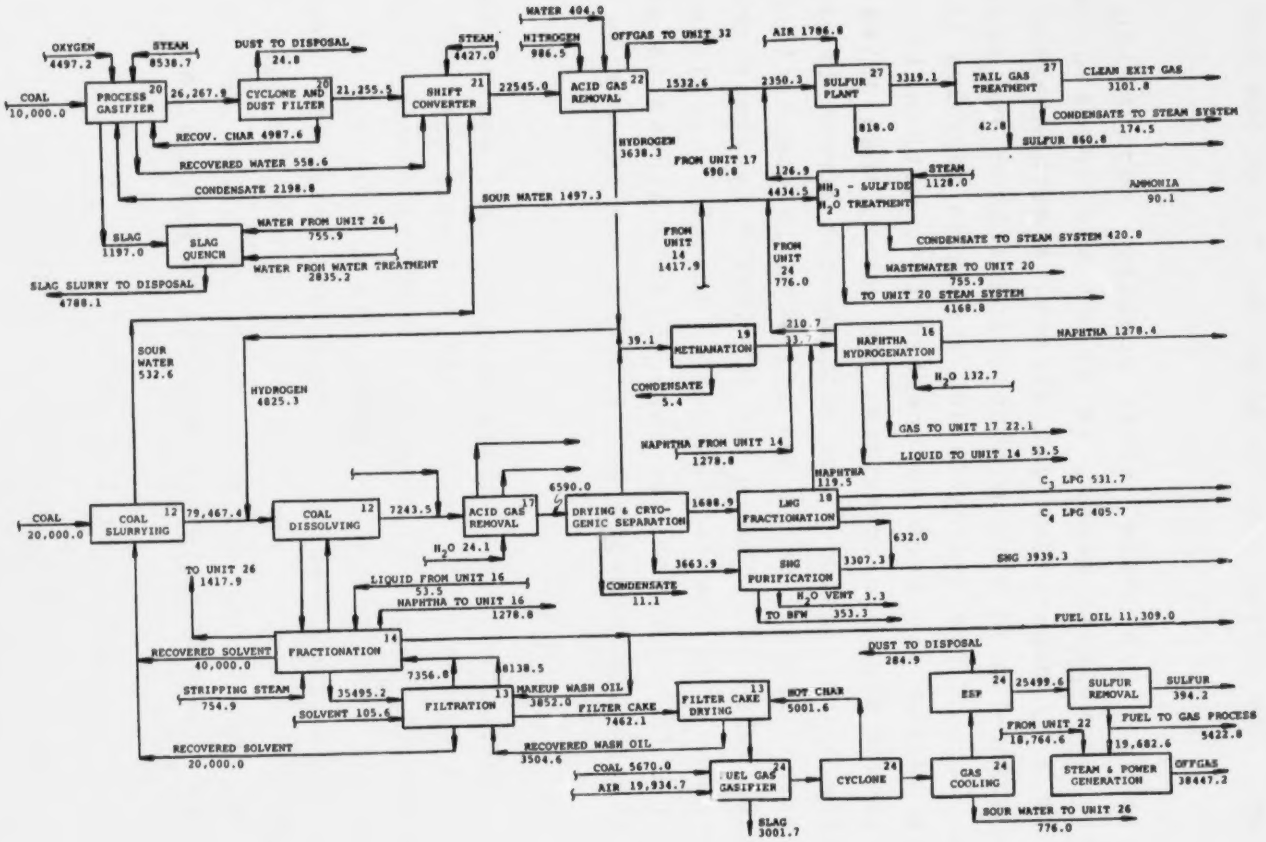
OXYGEN 4497.2
STEAM 8538.7
COAL 10,000.0
PROCESS GASIFIER 20
26,267.9
CYCLONE AND DUST FILTER 20
DUST TO DISPOSAL 24.8
21,255.5
RECOV. CHAR 4987.6
RECOVERED WATER 558.6
CONDENSATE 2198.8
SLAG 1197.0
SLAG QUENCH
WATER FROM UNIT 26 755.9
WATER FROM WATER TREATMENT 2835.2
SLAG SLURRY TO DISPOSAL 4788.1
STEAM 4427.0
SHIFT CONVERTER 21
22545.0
WATER 404.0
NITROGEN 986.5
ACID GAS REMOVAL 22
OFFGAS TO UNIT 32
1532.6
HYDROGEN 3638.3
FROM UNIT 17 690.8
SOUR WATER 1497.3
AIR 1786.8
2350.3
SULFUR PLANT 27
3319.1
TAIL GAS TREATMENT 27
CLEAN EXIT GAS 3101.8
CONDENSATE TO STEAM SYSTEM 174.5
818.0
42.8
SULFUR 860.8
126.9
4434.5
NH_3 - SULFIDE H_2O TREATMENT
STEAM 1128.0
AMMONIA 90.1
CONDENSATE TO STEAM SYSTEM 420.8
WASTEWATER TO UNIT 20 755.9
TO UNIT 20 STEAM SYSTEM 4168.8
FROM UNIT 14 1417.9
FROM UNIT 24 776.0
SOUR WATER 532.6
HYDROGEN 4825.3
39.1
METHANATION 19
CONDENSATE 5.4
210.7
33.7
NAPHTHA HYDROGENATION 16
NAPHTHA 1278.4
H_2O 132.7
GAS TO UNIT 17 22.1
LIQUID TO UNIT 14 53.5
NAPHTHA FROM UNIT 14 1278.8
COAL 20,000.0
COAL SLURRYING 12
79,467.4
COAL DISSOLVING 12
7243.5
ACID GAS REMOVAL 17
H_2O 24.1
6590.0
DRYING & CRYO-GENIC SEPARATION
CONDENSATE 11.1
1688.9
LNG FRACTIONATION 18
NAPHTHA 119.5
C_3 LPG 531.7
C_4 LPG 405.7
632.0
3663.9
SNG PURIFICATION
3307.3
SNG 3939.3
H_2O VENT 3.3
TO BFW 353.3
TO UNIT 26 1417.9
LIQUID FROM UNIT 16 53.5
NAPHTHA TO UNIT 16 1278.8
FRACTIONATION 14
FUEL OIL 11,309.0
RECOVERED SOLVENT 40,000.0
STRIPPING STEAM 754.9
35495.2
7356.8
8138.5
FILTRATION 13
SOLVENT 105.6
MAKEUP WASH OIL 3852.0
FILTER CAKE 7462.1
RECOVERED SOLVENT 20,000.0
RECOVERED WASH OIL 3504.6
FILTER CAKE DRYING 13
HOT CHAR 5001.6
COAL 5670.0
AIR 19,934.7
FUEL GAS GASIFIER 24
SLAG 3001.7
CYCLONE 24
GAS COOLING 24
SOUR WATER TO UNIT 26 776.0
ESP 24
DUST TO DISPOSAL 284.9
25499.6
SULFUR REMOVAL
SULFUR 394.2
FUEL TO GAS PROCESS 5422.8
FROM UNIT 22 18,764.6
19,682.6
STEAM & POWER GENERATION
OFFGAS 38447.2

Figure 5-4.- Coal liquefaction. SRC II block flow diagram and material balance.

carbon dioxide. An acid gas removal step separates carbon dioxide, hydrogen sulfide and other sulfur compounds from the gas stream, then the purified hydrogen is fed to the coal dissolver.

An air blown, slagging, suspension-type gasifier is fed with dried filter cake, recycled char and fresh coal. Slag residue is handled as previously described. Hot char is separated from the product gas stream and used to dry the filter cake. A sulfur removal step reduces the sulfur content of the gas sufficiently to permit use in steam generation and process furnaces.

Sour water collected from the process units is freed of oil in a separator. The oil is recycled to the fractionator in the coal liquefaction section. In a stripper column ammonia and H_2S are removed from the oil-free water and sent to ammonia recovery, while the cleaned water is sent to the process gasifier steam system. Ammonia is separated from H_2S in the Phosam (U.S. Steel proprietary) process by selective absorption into a monoammonium phosphate solution. The ammonia is subsequently stripped from the solution, dried, compressed, and liquefied as anhydrous ammonia. The overheads from the absorber are sent to sulfur recovery.

The Rectisol process, developed by the German Linde Company and Linde A.G., uses a proprietary solvent to absorb CO_2 and H_2S from the process gasifier product gas stream and then selectively desorbs the gases into an H_2S-rich stream that is sent to sulfur recovery and a CO_2-rich stream. The CO_2-rich stream, containing about 830 ppm of carbon monoxide and about 10 ppm of H_2O, is incinerated in the power generation boilers.

Sour gases from the coal dissolving unit and vent gases from the naphtha hydrotreater are combined and fed to a contactor where H_2S, COS and CO_2 are chemically absorbed in monoethanolamine

(MEA) solution. Clean gas becomes part of the SNG and LPG production. Regeneration of the amine solution yields an H_2S-rich stream that is sent to sulfur recovery.

SECTION 6

CURRENT ENVIRONMENTAL BACKGROUND: ENVIRONMENTAL REGULATIONS

INTRODUCTION

A survey was made of present and currently proposed environmental restrictions relevant to contaminants in the effluents, emissions,and wastes from coal conversion processes to serve as the measurement standard in evaluating available and developing control technology for such processes. The environmental restrictions set forth in Federal and state rules and regulations were reported together with selected international and regional regulations. The most stringent of the air and water regulations included herein have been summarized for convenience.

OBJECTIVES OF THE SURVEY

The prime objective of the survey was to assemble a single source reference document of applicable environmental regulations for use in considering both present control technology capabilities and necessary future technologies for controlling pollutants from the conversion of coal to gaseous or liquid fuel.

A second objective was to summarize the most stringent of the environmental regulations presented herein so that a single source of environmental requirements representing the most restrictive of present and proposed regulations would be available. A coal conversion facility built in the United States to meet the requirements in this most stringent summary would, by definition, meet the requirements of any individual state. The summary was by necessity limited primarily to those regulations of a quantitative (numerical) nature and did not include ordinances below the state jurisdictional level, since these were beyond the scope of the project. Special requirements introduced by individual states' permitting authorities were also beyond the scope of this project and were not included.

Another major objective was to provide an in-depth survey of the regulations of the selected states which had not been available previously to the extent presented in the survey. An example of the wide coverage of this survey is the inclusion of the U. S. EPA regulations applicable to Fluid Catalytic Cracking Units, Petroleum Refining Category, upon reasoning that giving a broad definition to Petroleum Refining, as some states do, makes such regulations potentially relevant to expected further on-site processing of coal liquefaction products.

BASIS FOR JURISDICTIONAL SELECTION

The first phase of the survey is concerned with federal and state environmental regulations. As such regulations are continually being amended they can only be reported current as of a given cut-off date. The cutoff date for the federal and state material in this report was 31 October 1977. The second phase supplements the first with a survey of regional and international regulations. Cutoff date for the second phase was 15 April 1978. On the premise that the first phase activity should be as broad as possible, it was decided that expanding the material considered relevant would be preferable to restricting it. Consequently, whenever it appeared that a particular standard or regulation might have at least some present or potential relevance, it was included in the survey. This approach was also advantageous with respect to use of the survey by project personnel as a source of guidelines to demonstrate the type and degree of restrictions placed on environmental contaminants.

The approach taken in the first phase was to collect, organize, review, and synopsize environmental laws, regulations, standards, and other restrictions of probable relevance. The coverage of this survey has intentionally been made as broad as possible to present the widest and most divergent restrictions in effect at both the federal and state levels. No local jurisdictional environmental requirements below the state level were included in the survey. As the commercial coal conversion facilities which are the underlying subject matter of this project are all yet to be built, only regulations pertaining to new facilities, as opposed to existing facilities, have been considered and included.

To make the initial collection and review of environmental factors as meaningful as possible, it was decided that the selection of the states to be included in this environmental survey would be based on the reported availability of coal deposits within the various states. This basis was chosen because economic factors favor sites near coal deposits for possible coal conversion plant locations. Accordingly, the environmental laws, regulations, and standards for 22 states were included with the federal restrictions in the first phase. The review of these state and federal requirements is broken down into the following three main areas:

Air Pollution Regulations
Water Pollution Standards
Solid Waste Requirements

A preliminary screening was carried out to determine most of the common contaminants normally present and to examine the state and federal regulations in view of these contaminants. The list of contaminants was extended as the work progressed.

It is noteworthy that, of all the states surveyed, only New Mexico has to date promulgated regulations which have specific applicability to fuel conversion facilities, and these regulations cover only the air pollution control area.

JURISDICTIONAL SELECTION

In addition to the federal environmental restrictions and guidelines included within this survey, regulations for the following states were selected based on the state's potential for future plant siting:

Alabama
Alaska
Colorado
Idaho
Illinois
Indiana
Kansas
Kentucky
Missouri
Montana
New Mexico
North Dakota
Ohio
Oklahoma
Pennsylvania
Tennessee
Texas
Utah
Virginia
Washington
West Virginia
Wyoming

In addition, the requirements as established by the U. S. Public Health Service Drinking Water Standards, 1962, and the Interim

Primary Drinking Water Regulations were synopsized and included in the survey.

Although California was one of the states initially selected for inclusion within the survey, because of time limitations and California's unique method of establishing environmental restrictions, California regulations are not being included. California also ranks quite low among the selected states in reported coal deposits. North Dakota, which has both sizeable coal deposits and projected conversion plants, was substituted for California. The review of federal and state standards was supplemented by a review of standards and guidelines established by the Delaware River Basin Commission, since the authority of this regional commission extends over geographical, rather than political, areas and therefore considers the area environment unconfined by artificial boundaries. It was found that the standards and guidelines adopted by the Susquehanna River Basin Commission are those of the states affected by the Commission.

Further consideration of the argument that environmental effects are not limited by political boundaries led to the inclusion in the survey of the standards and guidelines that have been established by Mexico and Canada. The Mexican regulations are federal actions, while in Canada both the Dominion and the provincial governments have enacted standards and guidelines. Therefore, Mexican federal standards, Dominion of Canada standards and guidelines, and the standards and guidelines of the provinces of Alberta and British Columbia became part of the survey. The two provinces were chosen because their boundaries are continguous with those of Montana, Idaho and Washington, where much of the U.S. western coal reserves are located.

Finally, the rules and guidelines established by U.S.-Canadian International Joint Commission were included in the survey, since these are primarily concerned with the Great Lakes and the St. Lawrence River areas and thus complete the regulatory coverage of the northern U.S. border.

METHOD OF INFORMATION ACQUISITION

The information required for the survey of environmental regulations was acquired in a series of steps:

> A review of the applicable Code of Federal Regulations (CFR) to determine the relevant federal environmental rules and regulations.
>
> A daily review of the Federal Register to determine applicable rules and regulations promulgated subsequent to the CFR edition, and additionally, any proposed rules, regulations, or notices which would be of concern.
>
> An initial request, prior to project inception, to the appropriate agencies of states selected requesting copies of their regulations. A request was also made that Pullman Kellogg be placed on mailing lists so as to be kept up-to-date on each state's activities in the environmental area.
>
> A subsequent request to all selected states upon project inception for their current regulations to ensure that the latest information would be on hand. Those states which had indicated that they were either revising their regulations or considering doing so were again contacted to determine the status and, where possible, draft regulations were obtained, reviewed and included within the survey.
>
> A review of Canadian regulations and guidelines, both

Dominion and provincial, in Pullman Kellogg's domestic and Canadian files for rules pertaining to the environment.

A review of the actions of the International Joint Commission of the United States and Canada, for relevant environmental rules.

A review of the appropriate Mexican regulations in the law library of the University of Texas in Austin.

A limited environmental literature survey was conducted through NTIS and selected EPA data base searches, and various technical periodicals were received and reviewed for useful information concerning federal and/or state environmental restrictions.

SPECIFIC ENVIRONMENTAL AREAS COVERED. COMMENTS

Air Emission Criteria

A review was conducted of the Federal EPA air emission regulations with designated emission source categories therein, of the selected state regulations, of the regional and the international regulations and guidelines. The potentially relevant air pollution standards were then synopsized for eventual inclusion in the survey report. As previously stated, the scope of this survey was purposely kept broad so as to provide the most comprehensive listing of existing and proposed regulations possible.

The following are comments generally applicable to a large majority of the jurisdictions surveyed:

A permit is required for construction, modification, or revision prior to commencement of the construction, modification, or revision contemplated.

Exceptions to the regulations are generally available for plant malfunction, startup, and shutdown so long as specified reporting requirements are complied with.

Dilution of any effluent or emission as a means of satisfying restrictions is prohibited.

Where there are several rules or standards applicable or more than one interpretation is possible, the most stringent should be applied.

In most jurisdictions, application for a variance from the established emission limits is possible, with discretionary approval authority in the jurisdiction's air pollution control agency.

Some states have specific geographical areas or air pollution control districts (sometimes heavily populated counties) which may have individual standards more stringent than the state-wide or "out-state" standards. No attempt was made to include these "localized" standards in this survey although a few of these regulations have been included for comparison.

Emission limitations applicable only to mobile sources were not considered as these types of regulations are not within the scope of the project.

The various regulations pertaining to monitoring of emissions were not included since these regulations are also beyond the scope of the review.

All the selected quantifiable standards as of the aforementioned cut-off dates were compared and the most stringent set of limitations was assembled; compliance with these limitations would presumably satisfy any criteria. The comparison of standards was undertaken as to numerical or quantified values only. Other regulations as found in the synopses would have to be considered in the design of proper control technology as well. These other regulations are primarily of a descriptive nature, sometimes of considerable length, and cannot readily be compared.

Water Effluent Limitations, Guidelines, and Standards

The Federal EPA effluent limitations and guidelines for specific point source categories and water quality standards of the selected states were reviewed and synopsized. As in the air pollution control area, the range of water quality criteria surveyed was purposely kept broad to provide the most comprehensive listing of standards possible.

A majority of the states and other regulatory bodies have established water quality standards which are applicable, for the most part, to existing receiving waters of the state. The primary state mechanism for controlling effluents into receiving waters is enforcement of the conditions imposed by a required discharge permit.

An analysis was made to determine the most stringent standards whenever a numerical comparison was possible, which standards would then presumably satisfy any jurisdictional criteria. Again, it should be emphasized that this was a comparison of quantifiable standards only, and other regulations would have to be considered.

Solid Waste Disposal Requirements

The Federal Guidelines and the selected states' solid waste disposal requirements were reviewed and these guidelines and standards were synopsized. The same policy as to scope and relevance of standards was used in this area as in the water and air regulation areas.

The majority of the solid waste disposal requirements are much less definitive, with regard to establishing design requirements, than those criteria established within the air and water regulatory areas. The regulations tend to establish requirements directed more toward the operation of a disposal facility than to the design, such as adequate rodent control and proper compaction and cover for solid waste. Even though the operational criteria should be considered beyond the scope of this survey, some of these standards are presented for certain selected jurisdictions as guidelines, since these types of criteria are essentially the same from area to area. Also, many of the requirements are applicable to public authorities, such as municipalities, in their solid waste collection and disposal activities.

It should be expected that the regulatory activity in this area and especially with respect to hazardous wastes will continue to increase as a result of the Solid Waste Disposal Act as amended by the Resource Conservation and Recovery Act of 1976, Title II Solid Waste Disposal (42 USC 6901 et seq.).

One provision generally common to the states reviewed allows for solid waste disposal on one's own property without a permit so long as no nuisance conditions are created.

Texas, one of the states surveyed, has issued Technical Guidelines for solid waste disposal and indicates that by following these guidelines all solid waste disposal requirements will be satisifed. These Technical Guidelines are available from the Texas Water Quality Board, which has responsibility in this area.

SUMMARY OF MOST STRINGENT WATER QUALITY STANDARDS

Notes:

1. The following compilation represents the most stringent criteria as established by the individual states, regions, and countries considered for this project.

2. It must be emphasized that this compilation represents an analysis based on numerical considerations only; compliance with these criteria should, in all probability, allow construction at any location. However, engineering design based on the following criteria may result in over design, and this should be considered for any cost data developed that are based on the criteria.

I. GENERAL CRITERIA FOR RECEIVING WATERS

A. The following minimum water quality conditions should be applicable to all receiving waters, and such waters should be:

1. Free from substances that will cause the formation of putrescent or objectionable sludge or bottom deposits.

2. Free from floating debris or other floating materials. (Alternate: Free from floating debris or other floating materials in amounts to be unsightly or deleterious.)

3. Free from substances producing color, or odor to the water.
(Alternate: Free from substances which produce

color or odor in amounts to be deleterious or to such degree as to create a nuisance.)

4. Free from substances in amounts which would impart an unpalatable flavor to fish.

5. Free from substances which would be harmful or toxic to human, animal, plant, or aquatic life.
   (Alternate: Free from substances in amounts which would be harmful or toxic to human, animal, plant, or aquatic life.)

6. Free from substances or conditions in concentrations which would produce undesirable aquatic life.
   (Alternate: Add to above, "Free from nutrients other deleterious substances attributable to sewage industrial wastes or other wastes.
   (Alternate: Add to above - in amounts which would affect public health or the desirability of the beneficial water use.)

7. Free from toxic substances, heated liquids, or any other deleterious substances attributable to sewage industrial wastes or other wastes.
   (Alternate: All to above - in amounts which would affect public health or the desirability of the beneficial water use.)

B. Acid Mine Drainage Control Measures (Applicable to coal processing)

1. Surface and ground water shall be diverted where practicable to prevent entry or reduce the flow into and through mine workings.

2. Refuse from the mining and processing of coal shall be handled and disposed of in a manner so as to minimize the discharge of acid mine drainage to streams.

3. Discharge of acid mine drainage to streams shall be regulated to equalize the flow of daily accumulation throughout a 24 hour period.

II. SPECIFIC WATER QUALITY STANDARDS - RECEIVING WATERS

A. The following specific water quality criteria should apply to all waters:

Substance or Condition	Limitation
pH (range)	7.0 to 8.8 (Br. Columbia)
Temperature	$\leq$ 1°C Rise (Canada-Federal) $\leq$ 60°F (Alaska and Washington) $\leq$ 85°F (North Dakota)
Dissolved Oxygen	$\geq$ 9.5 mg/l (Fresh water) $\geq$ 7.0 mg/l (Marine water) $\geq$ 5.0 mg/l (probable average)
Color	None $\leq$ 15 color Units (other criteria)
Turbidity	No Increase

Total Coliform Bacteria	_ 50/100 ml
Fecal Coliform Bacteria	≤ 10/100 mg (Domestic water supply) ≤ 200/100 ml (Probable average)
Settleable Solids	None (Essentially free) ≤ 200 mg/l - (Pennsylvania)
Dissolved Solids	≤ 100 mg/l (Br. Col., fresh water)
Oil and Grease	None ≤ 10 mg/l (Others)
Radioactivity	Gross beta - ≤ 100 pCi/l Strontium - ≤ 2 pCi/l Radium 226 - ≤ 1 pCi/l Alpha Emitters - 3 pCi/l
Odor and/or Taste	None ≤ 3 Threshold Odor Number (Probable average)
Total Dissolved Gas	≤ 110% of Saturation
Hardness	≤ 95 mg/l, max. 30 day avg. (Delaware River Basin Commission) (Delaware River Basin Commission)
Persistent Organic	Substantially absent

Persistent Organic Contaminants (harmful to human, animal, or aquatic life)	Substantially absent (North Dakota)
Toxic Substances	Persistant Toxicants - ≤ 1/2 of 96 hr TLM Non-Persistant Toxicants - ≤ 1/10 of 96 hr TLM
BODs (Deoxygenating Waste)	≤ 30 mg/l

B. The following chemical pollutants should not exceed the specified concentrations at any time:

Constituent	Concentration
Alkalinity	20-100 mg/l (Del. R. Basin, tidal waters)
Alkyl Benzene Sulfonate (ABS)	≤ 0.5 mg/l
Ammonia (as N)	≤ 0.02 mg/l (N. Dak; next value is ≤0.15 mg/l)
Arsenic	≤ 0.01 mg/l
Asbestos	Lowest Practicable Level (IJC*)
Barium	≤ 0.5 mg/l
Boron	≤ 1.0 mg/l
Cadmium	≤ 0.002 mg/l (≤0.01 Probable
Chloride	≤ 100 mg/l (≤ 250 probable average)
Chlorine, residual	≤ 0.002 mg/l (Proposed IJC)(Br. Col.: Below detectable limits)
Chromium (Hexavalent)	≤ 0.05 mg/l

Cobalt	≤ 1.0 mg/l
Copper	≤ 0.005 mg/l (Proposed IJC; 0.10 probable average)
Cyanide	≤ 0.005 mg/l
Fluoride	≤1.0 mg/l
H_2S, undissociated	≤0.002 mg/l (Proposed IJC)
Iron	≤ 0.3 mg/l
Lead	≤ 0.01 mg/l (Proposed IJC, Lake Superior; Ohio = ≤ 0.04)
Manganese	≤ 0.05 mg/l
Mercury	≤ 0.0002 mg/l (Proposed IJC)
Nickel	≤ 0.025 mg/l (Proposed IJC)
Nitrates ≤ 10 mg/l	
Phenols	≤ 0.001 mg/l
Phosphorus	≤ 0.05 mg/l
PCB (Polychlorinated biphenyl), total	≤ 0.00 mg/l
Selenium	≤ 0.005 mg/l (≤ 0.01 probable average)
Silver	≤ 0.0001 mg/l (Proposed IJC; ≤ 0.05 probable average)
Sulfate	≤ 250 mg/l
Uranyl Ion	≤ 5.0 mg/l
Zinc	≤ 0.03 mg/l (Proposed IJC)

III. EFFLUENT STANDARDS

(When Not Specified Differently by Discharge Permit)

Except as otherwise noted, compliance with the numerical standards should be determined on the basis of 24-hour composite samples, and no contaminant shall exceed five times the numerical standards at any time or in any one sample.

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*IJC = International Joint Commission of United States and Canada

A. No effluent shall contain the following:

1. Settleable Solids.
2. Floating debris.
3. Visible oil, grease, scum, or sludge solids.
4. Obvious color, odor and/or turbidity.
5. Fecal coliforms, concentration greater than 400/100 ml.

B. Additional contaminants, concentrations of which should not be exceeded in any effluent:

1.

Constituent	Concentration
Aluminum	$\leq$ 0.2 mg/l (Br. Col., one industry category)
Ammonia	$\leq$ 0.5 mg/l (Br. Col, tentative)
Antimony	$\leq$ 0.05 mg/l (Br. Col., one industry category)
Arsenic	$\leq$ 0.05 mg/l
Barium	$\leq$ 1.0 mg/l
Boron	$\leq$ 1.0 mg/l
Cadmium	$\leq$ 0.005 mg/l (Br. Col.)
Chlorate	50 mg/l (Br. Col, one industry category)
Chlorides	$\leq$ 250 mg/l
Chlorine, residual	0.2 mg/l (Br. Col., one industry category)
Chromium (Hexavalent)	$\leq$ 0.05 mg/l
Cobalt	0.1 mg/l (Br. Col., one

		industry category)
	Copper	≤ 0.05 mg/l (Br. Col.)
	Cyanide	≤ 0.02 mg/l
	Fluoride	≤ 1.0 mg/l
	Iron	≤ 0.3 mg/l
	Lead	≤ 0.05 mg/l (Br. Col.)
	Magnesium	150. mg/l (Br. Col., for fresh water; one industry category)
	Manganese	≤ 0.05 mg/l (Br. Col)
	Mercury	≤ 0.001 mg/l (Br. Col, tentative)
	Molybdenum	0.50 mg/l (Br. Col, one industry category)
	Nickel	≤ 0.2 mg/l (Br. Col)
	Nitrites (N)	10.0 mg/l (Br. Col., for several industry categories)
	Nitrogen	≤ 2.5 mg/l - April - Oct. ≤ 4.0 mg/l at other times
	Phenols	≤ 0.005 mg/l
	Phosphorus	≤ 1.0 mg/l
	Selenium	≤ 0.01 mg/l
	Silver	≤ 0.05 mg/l
	Sulfate	≤ 50 mg/l (Br. Col.)
	Sulfides and Mercaptans (S)	.011 mg/l (Br. Col., one industry category)
	Urea	1.0 mg/l (Br. Col., one industry category)
	Zinc (Ohio @ hardness < 80 mg/l as $CaCO_3$)	≤ 0.075 mg/l (Usual ≤ 0.1)
2.	BOD	≤ 30 mg/l (Deoxygenating Wastes)

3. COD	$\leq$ 125 mg/l
4. Temperature, max.	90°F (Br.Col., several industry categories)
5. Turbidity	$\leq$ 10 J.T.U. (Br. Col., several industry categories)
6. Solids: Total	$\leq$ 1,500 mg/l (Br. Col., several industry categories)
Dissolved (Total)	$\leq$ 1000 mg/l (Delaware R.B.C.)
Suspended	$\leq$ 25 mg/l (Canada-Federal)
7. Oil	$\leq$ 10 mg/l (Delaware R.B.C.)
8. Odor	250 (threshold number) (Delaware R.B.C.)
9. Persistent pesticides	Not to exceed 1/100 of TL50 value at 96 hours appropriate bioassay test (Delaware R.B.C.)
10. Dissolved oxygen (nontidal streams)	Not to reduce dissolved oxygen content of receiving water by more than 5% (Delaware R.B.C.)
11. Toxicity	50% max. mortality in 96 hours appropriate bioassay

	test with 1:1 dilution (Delaware R.B.C.)
12. pH*	6.5 to 8.5 (Br. Col. several industry categories)

IV. OTHER CRITERIA

A. Waste Treatment Ponds

Lagoons containing toxic substances or petroleum products must be lined. (Oklahoma)

B. Non-Degradation

Waters whose existing quality is better than the established standards shall not be lowered in quality.

C. Aesthetic values shall not be reduced by dissolved, suspended, floating,or submerged matter so as to affect water usage.

*The pH limitation should not be subject to averaging and should be met at all times.

SUMMARY OF MOST STRINGENT AIR QUALITY STANDARDS

Based on the Federal Standards, Selected States' Standards and the Regional and International Standards Covered in the Synopses

General Notes and Comments on Application and Use

1. The main objective of this analysis was to present the most stringent of the air standards covered in the synopses which could be compared numerically. Additionally, some of the shorter narrative or design specification types of regulation representing most stringent (or unique) requirements are also included. Topics generally having narrative type regulations of considerable length were not included. General fugitive dust emissions and storage and handling of organic materials and organic solvents were among those falling in this category.

2. New Mexico is the only state covered to have promulgated air regulations specifically for "Gasification Plants." All of these regulations have been included in appropriate sections in this compilation. A number of other regulations are included within this compilation which are unique to a certain jurisdiction and therefore automatically the most stringent. Many of these are for non-criteria contaminants, however.

3. Not all of the jurisdictions used the same basis for their standards for given contaminants and sources. Where possible conversion of all similar standards to a common base for

comparison was made but where this was impractical, the most stringent for each type of base remaining was presented.

4. The most general regulations with respect to geographical area or "out-state" regulations of the jurisdictions were synopsized and then compared. A few more stringent regulations for nonattainment areas (mainly heavily populated and industrialized counties) are shown in certain of the state regulations in addition to the "out-state" but were generally not included in the synopses.

5. Only regulations pertaining to new facilities were synopsized and compared as there are no existing commercial domestic fuel conversion plants of the type envisioned by this project.

6. Applicable Federal regulations found to be most stringent or as stringent as any jurisdiction covered are shown in this compilation in the appropriate place or category. Applicable but not most stringent Federal regulations are also shown but in parentheses for reference only. For easier reading the proposed Federal regulations for Stationary Gas Turbines have been placed together in one subsection of that title. They are newly proposed and the most stringent at present because none of the states covered have as yet promulgated regulations for such a source.

7. Prevention of Significant Deterioration (PSD) and Emission Offsets.
   The Clean Air Act Amendments of 1977 (enacted August 7, 1977) had considerable effect in the PSD area, which mainly provides the scheme for protecting areas with

air quality cleaner than the minimum national ambient air standards. Final regulation revisions for PSD were published in the Federal Register on June 19, 1978 at 43 FR 26380. State plans (SIP's) are required to reflect these requirements with revisions to be submitted to EPA by March 19, 1979. These regulations revise 40CFR Part 51 and Part 52 and are quite long and complex. As the publication date is after the April 15, 1978 cut-off date for source material for this project, a complete synopsis will not be made here of the PSD regulations as revised. However, a discussion of the effect and highlights of PSD follows due to the potential importance of these regulations relative to this project and the other regulations presented.

While PSD regulations apply primarily to areas meeting national ambient regulations for specified pollutants, Emission Offsets regulations apply primarily to dirty or nonattainment areas (areas not meeting ambient regulations). Because of the nature of the plants to be built relating to this project, it is assumed that attainment areas and thus PSD regulations are more relevant to the project. Two important basic requirements of Emission Offset regulations, where applicable, are: 1) lowest achievable emission rate (LAER) shall be attained, and 2) no net increase in emission shall result from an affected new or modified source. There are both significant similarities and differences in PSD and offset regulations.

A brief summary of the requirements set out in the PSD section of the 1977 Air Act Amendment (Section 165) follows:

> No major emitting facility (as defined in the Act) on which construction is commenced after August 7, 1977, may be constructed unless: 1) a permit has been issued setting forth emission limitations; 2) an air quality analysis has been conducted; 3) a public hearing has been held. (This is a new requirement which was not contained in earlier PSD regulations); 4) certain specified allowables (increments) are not exceeded; 5) best available control technology is applied; 6) the requirements for protection of pristine areas (Class I) have been met; 7) there has been an analysis of any air quality impacts projected for the area as a result of growth associated with the proposed facility; and 8) monitoring will be conducted to determine the effect of the facility's emissions on air quality.

PSD regulations at present apply to areas not exceeding National Ambient Air Quality Standards (NAAQS) for sulfur dioxide and particulates and establish allowable increases (incremental changes) for these pollutants in three area classifications above a defined baseline concentration. The allowable increases follow:

Pollutant	Maximum allowable increase (micrograms per cubic meter)
CLASS I	
Particulate matter:	
Annual geometric mean	5
24-hr maximum	10
Sulfur dioxide:	
Annual arithmetic mean	2
24-hr maximum	5
3-hr maximum	25

Pollutant	Maximum allowable Increase (micrograms per cubic meter)
CLASS II	
Particulate matter:	
Annual geometric mean	19
24-hr maximum	37
Sulfur dioxide:	
Annual arithmetic mean	20
24-hr maximum	91
3-hr maximum	512
CLASS III	
Particulate matter:	
Annual geometric mean	37
24-hr maximum	75
Sulfur dioxide:	
Annual arithmetic mean	40
24-hr maximum	182
3-hr maximum	700

Class I through III area classifications refer to geographical areas differentiated by the amount of incremental increases to be allowed in each. Class I increments permit only minor air quality deterioration; Class II increments, moderate deterioration; Class III increments, deterioration up to the secondary NAAQS.

Class I areas are often referred to as "pristine." Redesignation of lands from one classification to another by the states is allowed under some circumstances through specified procedures. Certain lands are now permanently in Class I. Class II increments have been changed and in some cases have become more restrictive; Class III increments now are specifically defined and procedures for reclassifying to Class III are more rigorous.

In the near future, PSD regulations will be extended to other pollutants for which NAAQS's are in effect (to be promulgated by August 8, 1979, and taking effect one year after promulgation).

It is now evident the required PSD air quality impact analysis will also apply in certain cases to nonattainment (dirty) areas. This is due to the possible effects of sources in nonattainment areas on air quality in clean areas. (The reverse can also be true so that the emission offset policy may have to be met by a clean area.)

A number of important and sometimes lengthy definitions are included in the current PSD regulations. For PSD purposes "major emitting source" under Section 169(1) includes 28 specified sources emitting, or having the potential to emit, 100 tons per year or more of any air pollutant. For those sources not specified only sources having emissions of more than 250 tons per year are subject to PSD requirements. "Baseline concentration" is defined as the ambient concentration level reflecting actual air quality as of August 7, 1977, minus any contribution from major stationary sources and major modifications on which construction commenced on or after January 6, 1975. Among other definitions are ones covering "major modifications," "potential to emit," "fugitive dust," "commence," "best available control technology," and "allowable emissions."

The two primary requirements that a PSD permit applicant must meet are:

1) Best available control technology, and

2) Not causing nor contributing to air pollution in excess of the maximum allowable increment or concentration for any pollutant more than one time per year.

Best available control technology is to be determined on a case-by-case basis, taking into account energy, environmental and economic impacts and other costs. At a minimum BACT must not result in emissions which would exceed the emissions allowed by new source performance standards under Section III or hazardous emission standards under Section 112 of the Air Act.

Of major concern is the likelihood of eventually consuming all available increments. The fuel conversions necessitated by the present energy situation in heavily industrialized areas will likely cause the sulfur dioxide and particulate increments to be exceeded, especially if all sources are counted against the increment. If offsetting reductions cannot be effected much of existing industry could be forced to curtail its operations and new sources could not be constructed.

The revised regulations follow the outline of the earlier regulations but, in general, are more comprehensive and restrictive.

8. The possible effects of certain general or catch-all provisions in the regulations covered must also be taken into account. Certain jurisdictions leave discretion in their air pollution agencies to lower specific numerical and other standards on the basis of "nuisance," for the "public welfare," because of specific health hazards, or where the application of best currently available control technology might reasonably dictate a more stringent standard. These would most often be applied on a case-by-case basis but could lower certain standards for an entire plant site under consideration. Because of the lack of specificity and wide variance in historical interpretation and application of such regulations among the jurisdictions, such regulations were not generally included in the stringency review. (Consideration of best currently available control technology with consideration of economic reasonableness and cost versus benefits is also a general requirement in the promulgation of regulations.) Typical examples of some of these general or catch-all provisions follow:

a) Nuisance - Interference with Enjoyment of Life and Property. Compliance with the regulations herein notwithstanding, should it be found after public hearing that any specific emission source is, will be, or tends to be significantly injurious to human health or welfare, animal or plant life, or property, or is or will be unreasonably interfering with the enjoyment of life and property of any inhabitant of the state, or will interfere with the attainment or maintenance of any national ambient air standards, alternate standards or orders may be issued to require additional abatement or control of certain emissions as deemed necessary to effect the purposes of the Kansas enabling act. (Kansas)

b) Air Quality Degradation Regulated. In areas of present high air quality where measured or estimated ambient levels of controllable pollutants are below the levels specified by applicable standards any emission of pollutant to the ambient air must be shown to result in pollution levels within applicable ambient air standards and will be prohibited in any case unless shown to be controlled to afford the highest efficiencies and the lowest discharge rates that are reasonable and practicable as specified in [subsection B.2]. (Utah)

c) Non-degradation. The significant and avoidable deterioration of air quality in any part of an area where presently existing air quality is equal to or better than that required by Ohio ambient air quality standards shall be prohibited. (Ohio)

d) More stringent requirements. A greater degree of control may be required to prevent a health hazard or a local nuisance because of the particular properties of a specific organic compound. Determination of a health hazard will be based upon such factors as threshold limit values, presence of carcinogens, and other accepted health indicators. (Indiana)

e) "Best available control technology" shall be determined on a case-by-case basis considering the following: (Virginia)

1) The process, fuels and raw material available and to be employed in the facility involved;

2) The engineering aspects of the application of various types of control techniques which have been adequately demonstrated;

3) Process and fuel changes;

4) The respective costs of application of all such control techniques, process changes, alternative fuels, etc.;

5) Any applicable emission standards; and

6) Location and siting considerations.

f) Best currently available control technology (BCACT). Air contaminant sources shall have installed and utilize the best currently available equipment and control technology for limiting emissions of gaseous air contaminants. (Tennessee)

g) Particulate Emissions - General Process Standards: Particulate non-attainment counties. In any county where one or more sources are emitting particulates at rates in conformity with applicable maximum emission rates and the ambient air quality standard for particulate matter is being exceeded, the Board shall set an appropriate emission standard for each source contributing to the particulate matter in the ambient air of the county at such value as the Board may deem necessary to achieve the desired air quality. (Tennessee)

h) Diluting and Concealing Emissions. The installation or use of any device, contrivance or operational schedule which, without resulting in reduction of the total amount of air contaminant released to the atmosphere, shall dilute or conceal an emission from a source is prohibited. (Wyoming)

9. Specific limitations and shortcomings in presenting this type of analysis in a conveniently brief manner are discussed in the last item of this section. The standards presented in this survey were condensed from generally lengthy rules and regulations. Numerical standards seldom stand alone and are generally clarified, modified, and limited by accompanying definitions, exceptions, calculation procedures and instructions, and other narrative forming the context in which they are found. The synopses herein attempted to retain the key points of such narrative but considerably less could be included in a reasonably brief stringency survey. Therefore, this survey may be used as a guide and convenient reference but not as a complete substitute for the synopses, the regulations, and other auxiliary sources such as court or agency opinions for in-depth regulatory applications.

10. Limitations and precautions in application. The following points indicate some of the most significient areas of potential difficulty or inconsistency with respect to the application of this most stringent air standards survey. The full implication of several of the points will probably only be apparent to those who now have or formerly have had extensive involvement in the practical application of air regulations.

a) Although fairly consistent, the rules for adding like facility rates or capacities within a given plant before calculating standards vary somewhat among the various jurisdictions and can lead to determination of different values for calculated standards. These rules are not always clearly spelled out in the written regulations and are often subject to or controlled by custom, board discretion, and court or agency rulings within a given jurisdiction. When necessary to add all units of a given source category within a plant before determination of the standard (probably the most common procedural method) smaller allowable emission values will generally result because percentage limitations on contaminants always decrease as capacity or stream rate increases where variable standards are specified.

b) As the most stringent regulations compiled are an artificial body of rules, no general system exists dictating controlling relationships between these regulations where conflict, overlapping, or the like might exist. Such a system does generally exist for any single jurisdiction and for the Federal and other jurisdictional regulations but these systems vary somewhat and thus would produce some differences upon application.

c) Some of the most stringent standards are presented by tables with standards at several specific stream rates or plant capacity levels. This was necessary where the jurisdictions used a mixture of constants and/or formulae to cover the standards over defined ranges, generally in terms of certain stream rates or other plant capacity indicating factors. To compare these for stringency it was necessary to solve any applicable formulae at some stream rate or plant capacity level within the relevant range. The fuel conversion plants under consideration are generally of very large throughputs and thus one or more points representing realistically high rates or capacities were among those selected in most cases. Conversely, some included units or streams in the plants could be rather small and so required other points to be determined representing rather small streams or capacities. A wide variety of plants and processes is within the project scope, and this introduces considerable potential variation in flow rates and sizes of specific units or facilities within the over-all plants.

11. Referring to the standards requiring tabular presentation as mentioned in 10.c) above, the most stringent standards could only reasonably be shown for a limited number of selected points. Interpolation between these selected points would almost never be proper because the points used could represent either constants or solutions to one or more formulae

at relevant capacities or rates. The synopses or the complete regulations would thus have to be referred to for intermediate standards evaluation, and in some cases care might have to be taken to determine which jurisdiction's regulation was most stringent at the new point in question.

I. Nitrogen Oxides (NO_x, expressed as NO_2).

A. Fuel Burning Equipment

Facilities ≥ 250 MM Btu/hr heat input*

Gas fired 0.2 lb/MM Btu (Federal, most
states) Liquid (oil) fired 0.3 lb/MM Btu (Federal,
most states) Solid (coal) fired 0.45 lb/MM Btu (N.M.;
Federal is 0.7 lb/MM
Btu)

Facilities <250 MM Btu/hr heat input

Any Fossil Fuel Best Available Control Technology
(Ohio)

Any size facility

Gas Fired 0.2 lb/MM Btu (Wyoming)
Solid (coal) fired 0.7 lb/MM Btu (Wyoming)

Facilities ≥ 1 MM Btu/hr heat input

Liquid (oil) fired 0.3 lb/MM Btu (Wyoming)

Facilities < 1 MM Btu/hr heat input

Liquid (oil) fired 0.6 lb/MM Btu (Wyoming)

Combined Fuel Firing (No Federal Standard but several states have a formula covering, Colorado's is shown below):

$$E = (0.2X+0.3Y+0.7Z)/(X+Y+Z)$$

Where: X is the % of total heat input from gaseous fossil fuel;
Y is the % of total heat input from liquid fossil fuel; and

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*Idaho incorporates the Federal standards here except that its Dept. determines on a case-by-case basis whether a stricter standard should be adopted after application of the best currently available control technology with reasons to be stated with the standard.

Z is the % of the total heat input from solid fossil fuel.

II. Visible Emissions

A. Processes

20% Opacity (#1 on Ringelmann Chart), (Kentucky)

B. General (Any Source)

20% Opacity (#1 on Ringelmann Chart), (Alabama and many states)

for ≥ 100,000 acfm flue gas rate: 15% Opacity (Texas)

C. Fuel Burning Equipment

10% Opacity (#0.5 on Ringelmann Chart), (West Virginia Federal - 20% Opacity)

D. Incinerators

10% Opacity (#0.5 on Ringelmann Chart), (Montana)

E. Coal Preparation Plants

Thermal Dryers - 20% Opacity, (Federal)

Pneumatic Coal Cleaning Equipment: 10% Opacity (Federal)

F. Petroleum Refineries

From fluid catalytic cracking unit catalyst regenerator or fluid catalytic cracking unit incinerator - waste heat boiler: 30% Opacity (Federal)

III. Particulates

A. Processes

1. Standards Based on Process Weight Rate

Process Weight Rate, lb/hr	Emission Standard, lb/hr
2.5 MM	54.2 (Alabama, others)
1.0 MM	46.8 (Alabama, others)
0.1 MM	20.5 (Illinois)
0.05 MM	14.2 (Illinois)

2. Standards Based on Stack Exhaust Gas Rate

Stack Exhaust Gas Rate, DSCFM	Concentration Standard, gr/DSCF
1 MM	0.02 (Pa.)
0.3 MM	0.02 (Pa.)
0.2 MM	0.03 (Pa.)
0.1 MM	0.04 (Pa.)

3. Process emitting 100 T or more of particulates annually based on 0 control (excluding combustion products of fuel oil, LPG, or natural gas).

   85% control of emissions (based on 0 control with source at maximum operating capacity). (Utah).

B. Petroleum Refineries

From fluid catalytic cracking unit catalyst regenerator or fluid catalytic cracking unit incinerator - waste heat boiler -

1.0 lb/1000 lb of coke burn-off (Federal), (incrementally higher emission rates are allowed for heat input attributable to auxiliary liquid or solid fossil fuels).

C. Gasification Plants - General (Certain boilers and coal briquet forming facilities within these plants are covered later in subsection III.E.).

Standard: 0.03 gr/scf exit gas

D. Fuel Burning Equipment

1. Standards Based on Heat Input Capacity.

(Fuels as designated)

Coal:	
>250 MM Btu/hr in	0.05 lb/MM Btu (H.H.V.) (New Mexico); and 0.02 lb/MM Btu (H.H.V.), for particulates 2 microns equivalent aerodynamic diameter or less (New Mexico)
<250 MM Btu/hr in	0.1 lb/MM Btu (Illinois)
Oil: >114 MM	0.005 (New Mexico)
<114 MM	0.1 (Illinois)
Gas: >2500 MM	0.1 (Texas)

Combinations Fuels: $E = S_s H_s + 0.10 H_1$ (Illinois)

where: S_s is applicable solid fuel particulate emission, lb/MM Btu actual heat input;

H_s is actual heat input from solid fuel, MM Btu/hr;

H_1 is actual heat input from liquid fuel, MM Btu/hr

(Any Fuel - Specific Fuels are not Designated)

Heat Input, Btu/hr	Standard, lb/MM Btu
10 MMM	0.1 (Federal, Okla.)
5 MMM	0.1 (Federal, several states)
500 MM	0.1 (Federal, several states)
50 MM	0.1 (Wyoming)
5 MM	0.4 (Ohio, other states)

2. Standards Based on Exhaust Gas Rate

Any fuel except coal or municipal waste: 0.05 gr/SCF (Alaska)

Coal or municipal waste: 0.1 gr/SCF (Alaska)

E. Gasification Plants (Designated Facilities and Fugitive Dust as per Respective Subsection Headings). (New Mexico)

1. Gas burning boilers (in conjunction with gasification plant)

   Standard: 0.03 lb/MM Btu heat input (L.H.V.)

2. Boilers firing more than one fuel (in conjunction with gasification plant).

   Formula for Standard:

   $$E_T = E_oQ_o+E_cQ_c+E_gQ_g$$

   where E_T is the total allowed emission in pounds per given period of time;

   E_o is the allowed emission from oil in lb/MM Btu's;

   E_c is the allowed emission from coal in lb/MM Btu's;

   E_g is the allowed emission from gas in lb/MM Btu's;

   Q_o is the heat released by the oil based on the higher heating value in Btu's per period of time;

   Q_c is the heat released by the coal based on the higher heating value in Btu's per period of time;

   Q_g is the heat released by the gas based on the lower heating value in Btu's per period of time.

Additionally, maximum emissions of particulates two microns or less (equivalent aerodynamic diameter) are limited by:

$$E_f = 0.40\ E_c\ (Q_o + Q_c + Q_g)$$

where E_f is maximum emissions in lb/given time period, and other terms are as defined in 2.a) above.

3. Coal briquet forming facilities

   Standard: 0.03 gr/SCF exit gas (with particulate emissions limited to stack outlets within technical feasibility).

4. Stack Design. All particulate emissions are to be through stacks at least ten diameters in length with adequate platforms and parts for sampling.

5. Fugitive Dust. No material shall be handled, transported, stored, or disposed of and no building or road shall be used, constructed, altered, or demolished without taking reasonable precautions to prevent particulates from becoming airborne.

F. Coal Preparation, Handling, and Mining

1. Standard for any thermal dryer: 0.031 gr/DSCF effluent gas (Federal)
2. Standard for any pneumatic coal cleaning equipment: 0.018 gr/DSCF effluent gas (Federal)
3. Coal preparation plants. All crushers, conveyors, screens, cleaners, hoppers, and chutes, which are designed for continuous transportation or preparation of coal shall be equipped with hoods, shields, or sprays where reasonably necessary to prevent airborne particulate matter. (New Mexico)
4. Coal mines-roads. Main coal haulage roads shall be sprayed or otherwise treated where reasonably necessary to prevent airborne particulate matter. (New Mexico)

G. Incinerators

1. Standards Based on Refuse Charge Rate

Refuse Charge Rate, lb/hr	Standard, gr/SCF Exhaust Gas (vol. corrected to 12% CO_2)
60,000	0.05 (Illinois)
5,000	0.08 (Federal, Illinois)
1,000	0.10 (Montana, others)
500	0.10 (Montana, others)
100	0.10 (Montana, others)

Refuse Charge Rate, lb/hr	Standard, lb/100 lb charged
500 (and higher rates)	0.10 (Ohio)
100	0.20 (Ohio)

2. Emissions with an excess of 100 ppm total carbonyls in the exhaust gases are prohibited. Operation shall only be during daylight hours unless permission for other operation is obtained from the Department. (Washington)

IV. Carbon Monoxide

A. General Sources

1. Emissions shall be limited so as to prevent ambient air standards for CO from being exceeded. Appropriate means are use of a direct flame afterburner or other Division approved means of equal effectiveness. (Wyoming)
2. All sources of CO shall control CO emissions by use of the best currently available control technology (BCACT, Ohio).

B. Petroleum Refineries or Processes

1. Any petroleum processing facility.
    a) A waste gas stream shall be burned in a direct flame afterburner or CO boiler so that a concentration of no more than 200 ppm (vol., corrected to 50% excess air) CO is emitted or shall be treated by other equivalent and Agency approved control technology. (Catalyst regenerators of fluidized catalytic converters equipped for in situ combustion of CO may emit CO concentrations up to 350 ppm corrected to 50% excess air. (Illinois) (The Federal standard and that of many states is 0.050% (vol.) in effluent gas.)
    b) Waste gas streams with CO from any catalyst regeneration of a petroleum cracking system, petroleum fluid coker, or any other petroleum process must be burned at 1,300°F for 0.3 sec or longer in a direct-flame afterburner or boiler with indicating pyrometer. (Alabama, Ohio)
2. Any petroleum process.
   Emissions shall be reduced by complete secondary combustion (93% removal of CO or more) of the waste gas stream. (Oklahoma)

C. Fuel Burning Equipment
1. Facilities with >10 MM Btu/hr actual heat input.
   200 ppm CO (corrected to 50% excess air). (Illinois)
2. Effluent streams with CO shall be burned in a direct-flame afterburner or boiler or controlled by other Board approved means. (Indiana)

D. Incinerators

1. Any incinerator
   500 ppm (corrected to 50% excess air). (Illinois)
2. Any effluent streams with CO shall be burned in a direct-flame afterburner or boiler or controlled by other Board approved means. (Indiana)

V. Odors

A. Odors from H_2S or Mercaptans.
Emissions containing H_2S or mercaptans shall be incinerated at 1,200°F or higher for at least 0.3 sec before discharge to the atmosphere or treated by alternate means shown to the Department to be at least as effective. (Pennsylvania)

B. Any source (some of the standards below represent similar requirements stated in slightly different ways by different states).

- Malodorous air from any source whatsoever, regardless of compliance with other odor standards [in these regulations] shall not be emitted such that any odor is detectable beyond the property line of such source. (Pennsylvania)
- Best available control technology as approved by the Board shall be used to limit odorous emissions from any odor emitting source. No odor, except for accidental or other infrequent emissions, that would be objectionable to a person of ordinary sensibility shall be emitted from a facility. (Virginia)
- The discharge of gases, vapors, or odors beyond the property line of an odor source so that a public nuisance is created is prohibited. (Montana)
- No odor shall be detectable from a sample taken at the property line of an odor source after dilution with up to seven volumes of odor free air as determined by the

Barneby-Cheney Co.'s centometer or equivalent and approved method. Effective odor control devices or systems shall be installed and operated such that odors in excess of the above are not created in areas adjacent to the source property line. (Wyoming; Missouri requires two separate tests not less than 15 minutes apart each hour.)

- Handling and storage. Odor producing materials shall be stored and handled so that accompanying odors do not create a public nuisance; accumulation of such quantities of these materials as to permit their escape or spillage shall be prohibited. (Montana)

C. Incinerators

1. Incinerators, including all associated equipment and grounds, shall be designed and operated in such manner as necessary to prevent emission of objectionable odors. (Ohio, Alabama, Missouri)

VI. Sulfur Dioxide and Sulfur

A. Fuel Burning Equipment - Sulfur Emissions Standards (Kansas)

$\geq$250 MM Btu/hr: 1.5 lb/MM Btu

B. Fuel Burning Equipment - SO_2 Standards

1. Specific fuels

Gas fired - SO_2 Standard: 0.16 lb/MM Btu L.H.V. (N.M. - Gasification Plants)

Residual oil fired* - SO_2 Standard: 440 ppm (vol.) SO_2 emissions concentration (Texas)

>115 MM Btu/hr: 0.34 lb/MM Btu (N.M.; Federal = 0.8 lb/MM Btu)

*(At very high rates the Kentucky SO_2 standard formula will be more stringent. This formula for oil fired equipment is: $Y = 7.7223X^{-04106}$.)

Distillate fuel oil fired - SO_2 Standard:
>250 MM Btu/hr: 0.3 lb/MM Btu (Illinois)
Coal fired - SO_2 Standard: >250 MM Btu/hr: 0.2 lb.MM Btu (Wyoming; Federal = 1.2 lb/MM Btu)
<250 MM Btu/hr: 1.2 lb/MM Btu (Oklahoma)

2. Any fuel
   SO_2 Standard: 1.0 lb/MM Btu (Ohio); also
   500 ppm (vol.)(Alaska)
3. Combination fuel fired. Several states have heat input dependent formulae for the SO_2 standard for combination fuels.

$$E = (0.8Y+1.2Z)/(Y+Z)$$

where: E is the maximum rate of emission, lb/MM Btu heat input (2 hr. avg.);
Y is the % of total heat input from liquid fuel; and
E is the % of total heat input from solid fuel

C. General Standards
1. SO_2 Standards:
   250 ppm (vol.) (British Columbia)
   1,000 lb/hr inexit gas (Mo.). Exceptions (Mo.): where SO_2 concentrations in ambient air at occupied places beyond emitting source premises don't exceed 0.25 ppm (vol.), 1 hr. avg., max. over once in a 4 day period; or
   0.07 ppm (vol.), 24 hr. avg., max. over once in a 90 day period.
2. Net SO_2 ground level concentrations (Texas)
   0.4 ppm, 30 min. avg., (allows exemptions when source meets Federal New Source Performance Standards, utilizes best available control technology, and doesn't cause or contribute to SO_2

primary and secondary ambient standards being exceeded in area. Several non-attainment counties have a 0.28 ppm, 30 min. average maximum).

D. Processes

1. Gasification Plants - S emissions (New Mexico): .008 lb/MM Btu heat input (HHV) in feed to plant
2. Sources other than fuel burning equipment and petroleum refineries - SO_2 Standards based on effluent concentrations.

   500 ppm (vol.) - current (Colo.)

   500 ppm (vol.) and emissions of not greater than 5T SO_2 per day from any process unit (applies only to SO_2 concentrations of 150 ppm (vol.) or more - effective on 1/1/85 and applies to new sources after 1/1/80 (Colorado)
3. Process with >250 MM Btu/hr heat input

   $E_p = 19.5P^{0.67}$ (Indiana)

   where $E_p = SO_2$ in lb/hr

   P = total process weight input capacity in T/hr

   at $P = 500$ T/hr, $E_p = 1254$ lb/hr
4. Ground level concentrations limits (if emitting >10 lb/hr SO_2) (Indiana):

   $$C_{max} = \frac{40\ S_p P^{0.75} n^{0.25}}{a h_s}$$

   where $C_{max.}$ = max. hourly ground level conc. with respect to distance and at the "critical wind speed for level terrain" resulting from the point source. $C_{max.}$ shall not exceed 900 ug/m^3 in areas where ambient air quality is better than applicable SO_2 secondary ambient air quality standards.

5. Regardless of the specific emission standard applicable in this regulation, emission sources shall utilize the

best available control technology as deemed appropriate by the Board. (Tennessee)

E. Sulfur Recovery Plants

1. SO_2 Standards - basis of lbs/lb S entering any size facility: .01 lb/lb S. (Ohio, Okla.)
2. SO_2 Standards - basis of lb/hr SO_2 allowed

   Texas formulae:

   >4000 SCFM effluent gas rate

   $E = 0.614\ q^{0.8042}$

   $\leq$ 4000 SCFM effluent gas rate

   $E = 123.4 + 0.091\ q$ (q = effluent gas rate; SCFM)

   Texas Standard at Rate Shown:

   @ 1 MM SCFM, E = 41,055 lb/hr
   @.237 MM SCFM, E = 6449 lb/hr
   @ 3 M SCFM, E = 396.4 lb/hr
3. S. recovery plants in conjunction with natural gas processing: 100 lb SO_2/hr, max. 2 hr. avg. (Okla.)
4. Also see VII D.5. - H_2S from sulfur recovery plants in conjunction with petroleum processing facilities. (New Mexico) That provision will generally be more stringent where applicable.

F. Sulfuric Acid Producing Facilities (Wyoming)

Processes producing H_2SO_4 by the contact method burning elemental sulfur, hydrogen sulfide, organic sulfides, mercaptans, or acid sludge shall limit SO_2 emissions in all effluent streams to:

not over 4 lb/T of acid produced, max. 2 hr. avg.

G. Petroleum Processing Facilities

1. Definition. "Plant processes" includes but is not limited to hydrogenation sweetening units, hydro-

cracking units, fuel burning equipment including flares and incinerators, sweetening system regeneration units, sour water strippers, and similar sulfur releasing systems. Catalyst cracking regeneration units other than hydrocracking units, boilers, or process heaters are not included if total emissions from them are less than 2.5T of S per day. (New Mexico)

2. Refinery plant processes (New Mexico)
   avg. S released per day >5T<30 T:
   .10 lb of S/lb S released in plant processes
   avg. S released per day $\geq$30 T:
   .02 lb of S/lb of S released in plant processes
3. Fuel-gas burning equipment (New Mexico)
   lb of S in effluent gas not to exceed a quantity equivalent to an S content of fuel gas entering of 10 gr/100 SCF of gas.
4. Fuel gas combustion devices in petroleum refineries (Federal).
   Fuel gas containing H_2S in excess of 0.10 gr/DSCF shall not be burned in any fuel gas combustion device. However, the combustion exhaust gases may alternately be treated so that the SO_2 emissions control is the equivalent with respect to SO_2 of compliance with this H_2S content regulation.
5. Alabama regulations, which cover petroleum facilities handling natural or refinery (process) gas containing more than 0.10 gr H_2S/SCF of gas could be more stringent in certain isolated cases. These

regulations have an increment correction factor (additional SO_2 emission allowed) dependent on mol % H_2S in the dry acid gas up to 60 mol % H_2S. Alabama also has a requirement that a demonstration be made that the SO_2 emitted will not cause or contribute to non-attainment of any primary or secondary ambient air standards. For reference only the basic uncorrected Alabama regulations are as follows:

>10 Lt ≤ 50 LT available S per day: 560 lb/hr
>50 Lt $\leq$ 100 LT available S per day: 0.10 lbs/ SO_2/lb of S processed
>100 LT available S per day: 0.08 lbs SO_2/lb S processed

H. Standards for Sulfur Content of Fuels (as S). (Montana)
   1. Liquid or Solid Fuel
      a) Max. S in Fuel: 1 lb/MM Btu fired
   2. Gaseous Fuels
      a) Max. S in Fuel (calculated as H_2S): 50 gr/100 SCF of fuel
   3. Exceptions and exemptions to the standards in 1. and 2. are listed.

VII. Hydrogen Sulfide

A. Any source-general (Texas)
   1. Max. net ground level concentration.
      a) Where residential, business or commercial property downwind of H_2S source is affected
         0.08 ppm, 30 min. avg.
      b) Where H_2S source affects only downwind property used for other than the purposes listed in 1.a) above (such as industrial, vacant tracts, or range lands not normally inhabited.)
         0.12 ppm, 30 min. avg.

B. Gasification Plants (New Mexico)

1. 100 ppm (vol.) in effluent gas, to any combination of H_2S, carbon disulfide, and carbon oxysulfide, and
2. 10 ppm (vol.) in effluent gas, max. H_2S component in combined effluent gas.

C. Processes

1. H_2S emission rate (or rate to any combustion device) (Kansas):

   10 gr/100 ft^3 of gas emitted (or fed to combustion device); except combustion of fuels where the gaseous

   products are used as other process raw materials.
2. Limits on H_2S in ambient air in inhabited areas beyond the source premises where concentrations are attributable to such source. (Missouri)

   0.03 ppm (vol.), 30 min. avg., not more than twice in any 5 consecutive days

   0.05 ppm (vol.), 30 min. avg., not more than twice per year
3. Effluent gas from H_2S process sources shall be vented, incinerated, flared or otherwise disposed of so that ambient H_2S and SO_2 standards are not. exceeded. (Wyoming)

D. Petroleum Processing Facilities

1. Fuel combustion devices (Federal):

   Fuel gas containing H_2S in excess of 0.10 gr/DSCF gas shall not be burned. The exhaust gases may alternatively be treated so that equivalent SO_2 emission control is obtained upon such showing to the satisfaction of the EPA Administrator.

   Exceptions. Flaring of process upset gas or of the process or fuel gas from relief valve leakage is exempt from the above.

2. Maximum H_2S concentration in refinery process gases emitted or combusted: 150 ppm (Alabama)

3. Ground level H_2S concentration beyond source premises for facilities handling natural or refinery (process) gas containing more than 0.10 gr H_2S/SCF. (Alabama)
   Stream shall be burned to maintain a 20 ppb concentration, averaged over a 30 min. period. (Determination of such concentration from waste gas or emergency flaring to assume only 75% of heat of combustion is used to heat products of combustion).

4. H_2S emissions shall be controlled by removal of H_2S from exhaust gas or H_2S oxidation to SO_2 in a system insuring complete oxidation of H_2S at all times. H_2S limits in either type of control system shall be: (Oklahoma)
   0.3 lb/hr of H_2S, 2 hr. avg.; and 95% removal of H_2S in exhaust gas Any oxidation system shall utilize a stack at least 50 feet in height. Such system shall not be allowed to emit over 100 lb/hr of SO_x (expressed as SO_2 2 hr. avg.) unless there is a prior removal step meeting Oklahoma SO_x limitations.

5. H_2S from petroleum processing facilities including sulfur recovery plants in conjunction with such facilities (New Mexico). Either:
   a) 10 ppm (vol.) max. in effluent gas; or
   b) the effluent gas shall be passed through suitable equipment to oxidize the H_2S to SO_2
   c) Flares which may flare gas with over 10 ppm of H_2S shall utilize alarms to signal non-combustion of the gas.

6. Claus Sulfur Recovery Plants. No discharge shall be allowed of any gases to the atmosphere containing in excess of:
   a) 0.025% (vol.) of SO_2 at 0.0% O_2, dry basis, where emissions are controlled by an oxidation control system, or a reduction control system followed by incineration, or
   b) 0.030% (vol.) of reduced sulfur compounds and 0.001% (vol.) of H_2S calculated as SO_2, at 0.0% O_2 on a dry basis, if emissions are controlled by a reduction control system not followed by incineration.

VIII. Sulfur Content of Fuels

A. Maximum Sulfur Content (wt % S or as noted)

Any fuel oil	1.5% (Utah)
Distillate fuel oil	
#1	0.3% (Idaho)
#2	0.5% (Idaho)
Coal (Solid)	1.0% (Idaho) 1 lb S/MM Btu input (Montana)*
Gaseous fuel	50 gr (as H_2S)/100 SCF of fuel in (Montana)*

IX. H_2SO_4, Sulfuric Acid Mist, SO_3

A. Emissions of H_2SO_4 or SO_3 (or combination)

1. 35 mg (as H_2SO_4)/m^3 of effluent gas (Missouri)

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*Montana allows higher sulfur content fuels with proper approval where such fuels are mixed with lower sulfur-containing fuels so that the mixture doesn't exceed the standard. Montana also allows SO_2 emission control in the alternative if such control will be equivalent in terms of sulfur emitted (in lb/hr).

2. Process sources not including sulfuric acid manufacturing sources (Illinois)
   - equivalent sulfuric acid usage $<1,300$ tons per year (100% acid basis)
     - 0.10 lb in any 1 hr period
   - equivalent sulfuric acid usage $\geq$ 1,300 tons per year (100% acid basis)
     - 0.50 lb /T of H_2SO_4 used

B. Concentration of H_2SO_4 or SO_3 (or combination) in Ambient Air in Inhabited Areas beyond Source Premises (Missouri)
   - 0.03 mg (as H_2SO_4)/m^3, 30 min avg., not over once in 48 hrs.
   - 0.01 mg (as H_2SO_4)/m^3, 24 hr avg., not over once in 90 days
   - 100 ug/m^3 of air (std) at any time

C. Net Ground Level Concentration (Texas)
   - 15 ug/m^3 of air (std), 24 hr avg.
   - 50 ug/m^3 of air (std), 1 hr avg., measured more than once in any 24 hr period.

X. Other Miscellaneous Sulfur Compounds

A. Mercaptans

1. Petroleum processing facilities:
   Emissions of mercaptans shall be either: not greater than 0.25 lb/hr (total mercaptans), or controlled by passing through a combustion device which will achieve complete combustion or any other equally efficient device for control of mercaptans. (New Mexico)

XI. Gasification Plants - Other Contaminants

A. HCN Standard.
   - 10 ppm (vol.) in effluent gas (New Mexico)

B. NH_3 Standard.

25 ppm (vol.) in effluent gas (New Mexico)

Additionally any stationary tanks holding NH_3 shall be:

A pressure tank capable of maintaining working pressures sufficient to prevent loss of NH_3 to the atmosphere, or

Equipped with other equally effective control equipment to prevent loss of NH_3 to the atmosphere. (New Mexico)

C. Hydrogen Chloride/Hydrochloric Acid Standard.

5 ppm (vol.), any combination of hydrogen chloride and hydrochloric acid in effluent gas. (New Mexico)

XII. Hazardous Air Pollutants

A. Definitions. "Hazardous air pollutants" means an air pollutant to which no ambient air quality standard is applicable and which in the judgement of the Administrator causes or contributes to air pollution which may reasonably be anticipated to result in an increase in mortality or an increase in serious irreversible or incapacitating reversible illness. (Federal)

B. Mercury (Federal)

1. Definition. "Mercury" means the element mercury, excluding any associated elements, and includes mercury in particulate, vapors, aerosols, and compounds.
2. Emission standard. Emissions from sludge incineration plants, sludge plants, or combinations of these that process wastewater treatment plant sludges shall not exceed 7.05 pounds of mercury per 24-hour period.

C. Beryllium

1. Federal Standard.

   Affected facilities. Extraction plants, ceramic plants, foundries, incinerators, and propellant plants which process beryllium ore, beryllium, beryllium oxide, beryllium alloys, or beryllium-containing waste.

   Emission Standard. Emissions from stationary sources subject to this provision shall not exceed 10 grams of beryllium over a 24-hour period.

2. Texas Concentration Standard

   0.01 ug/m^3, 24 hr avg. (To be measured by the difference between upwind and downwind concentration levels for the source premises, or by stack sampling, calculated to a downwind concentration (details in appendices of Texas Air Regulations.)

D. Hazardous Pollutants - General

1. The utmost consideration shall be given to the potential harmful effects of and effective control methods for discharages to the open atmosphere of hazardous matters including, but not limited to, antimony, arsenic, asbestos, beryllium, bismuth, lead, mercury, silica, tin, and compounds of such materials. Evaluation of these sources and the control methods designed and proposed will be made on a case-by-case basis by the Department. (Kentucky)

E. Two of the selected states, Virginia and Colorado, have chosen to incorporate into their hazardous contaminants category extensive lists of elements and compounds from sources outside the EPA which are aimed primarily at protecting humans from excessive exposure levels of these contaminants. Virginia has incorporated the OSHA air contaminants list in 29CFR 1910.1000 while Colorado has incorporated the ACGIH's published list of "threshold Limit Values" (TLV's) as adopted at the ACGIH 35th Annual Meeting in May 1973. The method each of these states utilizes to determine standards and to enforce such levels for these substances is described in their respective synopses, which also include copies of the complete lists. No attempt has been made to identify relevant compounds on these lists for this project as this falls in line with other work projected for the future work.

XIII. Other Non-Federal Contaminant Regulations Unique to One or Only a Few States

A. Ice Fog (Alaska)

1. Any person proposing to build or operate an industrial process, fuel burning equipment, or an incinerator in an area of potential ice fog may be required to reduce water emissions and to obtain an operating permit.

B. General Gaseous Emission Standards

1. Non-process (Tennessee)
    - a) Definitions. "Air contaminant source" for subsection B.1. means any and all sources of emission of air contaminants, whether privately or publicly owned. Without limiting the generality of the foregoing, this term includes all types of business, commercial, and industrial plants, works, shops, and stores, and heating and power plants and stations, buildings and other structures of all types,

incinerators of all types (indoor and outdoor), refuse dumps and piles, and all stack and other chimney outlets from any of the foregoing.

b) Standard.Air contaminant sources shall install and utilize the best currently available equipment and control technology.

2. Process (Tennessee)
Gaseous air contaminant sources shall utilize equipment and technology deemed reasonable and proper by the Board for control of emissions of such contaminants.

C. Mineral Acids - Nitric Acid Mist or Vapor, Hydrochloric Acid Mist,or Vapor

1. Allowable stack gas concentration from any stationary sources.

Nitric Acid Mist and/or Vapor: 70 mg/DSCM
(West Virginia)
Hydrochloric Acid Mist and/or Vapor: 210 mg/DSCM
(West Virginia)

D. Fluorides

1. Inorganic fluoride compounds
6 ppb (vol., 3 hr avg., expressed as HF), (Texas)

2. Fluorine, fluorides (Idaho)
It is prohibited to discharge such quantities (in combination with all other sources of fluorine and fluorides, both natural and man-made) that the total fluoride content in vegetation for feed or forage resulting from contact with the ambient air exceeds:

a) 400 ppm (dry) - annual arithematic mean

b) 60 ppm (dry) - monthly conc. for 2 consecutive months

c) 80 ppm (dry) - monthly conc. never to be exceeded

3. Phosphate, phosphorite, or phosphorous processing equipment and other fluoride processing or treating equipment. (Montana)

   a) Any phosphate rock or phosphorite or phosphoric acid processing equipment, or equipment processing fluorides enriched wastewater or fluorides in gaseous or particulate form or combinations.

      0.3 lb/T of P_2O_5 introduced (fluoride releasing processes)

   b) Pond emissions. Any fluoride emissions from storage ponds, settling basins, ditches, liquid holding, or conveying tan.. or device associated with facilities in 3.a) above.

      108 ug/cm^2/28 days (calcium formate method)

XIV. Stationary Gas Turbines (Federal - Standards of Performance for Stationary Sources)

A. Affected facilities: This subpart shall be applicable to all stationary gas turbines with a heat input at peak load equal to or greater than 10.7 gigajoules per hour, based on the lower heating value of the fuel fired.

B. Definitions

   1. "Stationary gas turbine" means any simple cycle gas turbine, regenerative cycle gas turbine, or any gas turbine portion of a combined cycle steam/electric generating system that is not self-propelled. It may, however, be mounted on a vehicle for portability.

2. "Simple cycle gas turbine" means any stationary gas turbine which does not recover heat from the gas turbine exhaust gases to preheat the inlet combustion air to the gas turbine, or which does not recover heat from the gas turbine exhaust gases to heat water or generate steam.

3. "Regenerative cycle gas turbine" means any stationary gas turbine which recovers heat from the gas turbine exhaust gases to preheat the inlet combustion air to the gas turbine.

4. "Combined cycle gas turbine" means any stationary gas turbine which recovers heat from the gas turbine exhaust gases to heat water or generate steam.

C. Emission Standards for Nitrogen Oxides (NO_x)

1. Gas turbines with heat rate at peak load $\geq$ 14.4 kilojoules per watt hr (lower heating value of fuel).

   NO_x emissions, E, in exit gases not to exceed:

   $E = 0.0075 + F$

   where:

   E = maximum NO_x emissions in % by volume (at 15% O_2 and on dry basis)

   F = NO_x emission allowance for fuel bound nitrogen as defined in subpart C.3. below.

2. Gas turbines with heat rate at peak load < 14.4 kilojoules per watt hr (lower heating value of fuel).

   NO_x emissions, E, in exit gases not to exceed:

   $$E = 0.0075 \frac{14.4}{Y} + F$$

where:

E = maximum NO_x emissions in % by volume (at 15% O_2 and on dry basis)

Y = manufacturer's rated heat duty at peak load in kilojoules per watt hour.

F = NO_x emission allowance for fuel bound nitrogen as defined in C.3. below.

3. The term F shall be defined according to the nitrogen content of the fuel as follows:

Fuel Bound N* (% by Wt)	F (NO_x % by vol.)
$N \leq 0.015$	0
$0.015 < N \leq 0.1$	0.04 (N)
$0.1 < N \leq 0.25$	0.004 + 0.0067 (N-Oil)
$N < 0.25$	0.005

*N = Weight % nitrogen in the fuel

4. Exemptions from NO_x Emission Standards
   a) Stationary gas turbines with a heat input at peak load of 107.2 gigajoules per hour (100 MM Btu/hr) or less (lower heating value of fuel) - exempt for not more than 5 years from proposal date of these rules.
   b) Stationary gas turbines using water or steam injection for control of NO_x emissions - exempt when ice fog is deemed a traffic hazard by turbine owner or operator.
   c) Emergency standby gas turbines, military gas turbines other than at garrison facilities, and fire-fighting gas turbines.

D. Emission Standards for Sulfur Dioxide (SO_2)

1. Maximum Emission Rate in Exit Gases

   0.015% SO_2 (by vol., at 15% O_2 and on dry basis)

2. Use of fuel sulfur content in determination of compliance with subpart D.1. This method may be utilized in the alternative with a maximum sulfur content of 0.8% by weight in any fuel burned by a gas turbine under such circumstances.

SECTION 7

DEVELOPMENT OF ENVIRONMENTAL OBJECTIVES

The compilation of present and proposed environmental standards and guidelines for federal, state, regional, and international jurisdictions, and the summarization of these regulations into a listing of most stringent standards, provides a set of criteria against which the engineering designs for control of effluents, emissions and wastes from proposed coal conversion plants may be judged. The statement may be made that a coal conversion facility designed to meet the most stringent standards could be built anywhere in the United States, Canada, or Mexico.

By definition, the most stringent standards are summarizations of existing legislation and reflect the opinions and best judgement of the legislators at the time of enactment. As more knowledge accumulates concerning the great variety of organic and inorganic compounds that may be formed during conversion of coal, and more data are gathered on the effects of these compounds on the environment and its inhabitants, the possibilities arise that existing standards may be right, or too stringent or not sufficiently stringent. There is the additional possibility that the existing standards are not sufficiently comprehensive and, in the future, should be extended to include more materials.

Assistance in evaluating present standards against present needs, and in indicating possible future goals for legislative improvement, is provided by EPA through the establishment of Multimedia

Environmental Goals (MEGs). The MEGs are estimates of desirable ambient and emission levels of control and, as such, are an integral part of EPA's environmental assessment approach. The program to assemble a master list of chemical contaminants, complex effluents and mixtures and then to develop a methodology to establish meaningful values that will serve as MEGs is described at length in "Multimedia Environmental Goals for Environmental Assessment" by J. G. Cleland and G. L. Kingsbury, EPA-600/7-77-136a(Vol. 1) and -136b(Vol. 2), November 1977.

COMPARISON OF MOST STRINGENT REGULATIONS WITH MEG CRITERIA

The stringency criteria developed from the synopses of air and water regulations were compared with the criteria for like contaminant substances shown in MEG charts in the MEG report previously noted. Four categories of concentrations were compared: Water Effluents, Water Ambient Bodies, Air Effluents, and Air Ambient Bodies.

Values were not always available for all of the categories for a given substance. Further, since many important regulations relevant to the project were stated only in terms of allowable rates, and were thus not comparable to MEG concentration data, their provisions were not included in the comparison. A total of 43 relevant substances were compared, with the following results:

	More Stringent		Same Value	Total
	MEG	Regulations	For Both	Comparable
Water Effluent	7	12	14	33
Water Ambient	8	4	8	20
Air Effluent	12	1	0	13
Air Ambient	1	1	3	5
	28	18	25	71

The comparison for emissions/effluents is shown in TABLE 7-1. The comparison for ambient bodies is shown in TABLE 7-2.

The primary purpose of this comparison and study was to determine the relative level of stringency represented by the most stringent of the regulations synopsized as compared with the most stringent of a body of criteria known to be, as MEG criteria are, primarily based on actual effects on human welfare and/or the environment (such criteria as might be expected to be utilized by major regulatory agencies in the formulation of new anti-pollution standards or in updating existing standards). With the criteria available in the MEG charts it was possible to compare criteria for contaminants developed for air and water effluent streams and ambient (receiving) air and water bodies.

Similar solid waste media comparisons were not made because the solid waste synopses contained few of the numerical standards that are needed for such a quantitative analysis or comparison.

The values used in the comparisons for most stringent air and water standards came from the summaries of the most stringent air and water quality standards in this report and from the report section covering ambient air quality standards.

The methodology described in "Multimedia Environmental Goals for Environmental Assessment" makes use of charts for emission level goals and ambient level goals as shown in Figure 7-1. Both goals were considered in the comparision with most stringent standards. Emission level goals are those associated with effluent or discharge streams from point sources and/or with fugitive emissions. These goals may be based on technology factors or ambient factors; however, technology based emission criteria were not addressed by the MEG report and, therefore, ambient factors only served as the MEG criteria used in comparisons of emission level goals.

TABLE 7-1. COMPARISON OF MOST STRINGENT STANDARDS CRITERIA WITH MEG CRITERIA FOR EMISSIONS/EFFLUENTS

Comparision of Most Stringent Water and Air Effluent Concentrations with Multimedia Environmental Goals (MEGs) for Emissions. MEGs Based on Ambient Factors Rather Than Best Technology Available.

	Maximum Effluent Concentration			
	In Water, mg/l		In Air, mg/m3*	
Pollutant	MEG MATE**	Most Stringent	MEG MATE**	Most Stringent
Alkyl benzene sulfonate (ABS)	-	-	-	-
Aluminum	0.073	0.2	0.0126	-
Ammonia	0.05	0.5	0.35(0.49)	(25)
Antimony	0.2	0.05	0.5	-
Arsenic	0.05	0.05	0.002	6.9
Barium	2.5	1.0	0.5	-
Beryllium	0.03	-	0.002	0.00001 (24 hr)
Boron	25	1.0	3.1	-
Cadmium	0.001	0.005	0.01	6.9
Carbon Monoxide	0.06	-	40(35)	(200)(93% removal)
Chlorate	-	50	-	-
Chloride	-	250	-	-
Chlorine, residual	-	0.2	-	-
Chromium	0.25	0.05	0.001	-
Cobalt	0.25	0.1	0.05	7.0
Copper	0.05	0.05	0.2	6.0
Cyanide(as HCN)	0.025	0.02	11.0	-
Fluoride (as HF)	-	1.0	-	0.006 (3 hr)
Hydrogen chloride, HCl (acid), HCl (mist)	-	250	-	210
Hydrogen sulfide	0.01	-	15(10)	(100)
Iron	-	0.3	-	-
Lead	0.05	0.05	0.15	6.9
Magnesium	87.0	150	6.0	-
Manganese	0.1	0.05	5.0	-
Mercury	0.01	0.001	0.01	6.9
Molybdenum	7.0	0.50	5.0	-

TABLE 7-1. COMPARISON OF MOST STRINGENT STANDARDS CRITERIA WITH MEG CRITERIA FOR EMISSIONS/EFFLUENTS (CONT)

Pollutant	Maximum Effluent Concentration			
	In Water, mg/l		In Air, mg/m^3*	
	MEG MATE**	Most Stringent	MEG MATE**	Most Stringent
Nickel	0.01	0.20	0.015	-
Nitrates	-	-	-	-
Nitrites(as N)	-	10.0	-	-
Nitrogen	-	2.5	-	-
Nitrogen oxides	140	-	9.0	-
Ozone	-	-	0.01	-
Phenols	0.005	0.005	19.0(5)	100(26)
Phosphorus	0.0005	1.0	0.1	-
Polychlorinated biphenyl (PCB), total	0.000001	-	0.5	-
Selenium	0.025	0.01	0.2	-
Silver	0.005	0.05	0.01	-
Sulfate	-	50	-	-
Sulfides and mercaptans (as S)	15.0	0.011	1.0(0.5)	-
Sulfur	-	-	-	-
Sulfur dioxide	200	-	13	250
Urea	-	1.0	-	-
Zinc	0.1	0.075	4.0	6.9

* Concentrations in parentheses are ppm(v)

**MATE = Minimum Acute Toxicity Effluent value from the applicable MEG charts. The value shown is the more stringent of health or ecological effects criteria.

TABLE 7-2. COMPARISON OF MOST STRINGENT STANDARDS CRITERIA WITH MEG CRITERIA FOR AMBIENT BODIES

Comparison of Most Stringent Water and Air Criteria for Ambient (Receiving) Bodies With Multimedia Environmental Goals (MEGs) for Ambient Bodies

	Maximum Effluent Concentration			
	In Water, mg/l		In Air, mg/m^3*	
Pollutant	MEG MATE**	Most Stringent	MEG MATE**	Most Stringent
Alkyl benzene sulfonate (ABS)	-	0.5	-	-
Aluminum	1.0	-	5.2	-
Ammonia	0.010	0.15	0.035	-
Antimony	0.007	-	0.0012	-
Arsenic	0.01	0.01	0.000005	-
Barium	0.5	0.5	0.001	-
Beryllium	0.000075	-	0.00001 (30 day)	0.00001 (30 day)
Boron	0.043	1.0	0.074	-
Cadmium	0.0004 (fresh) 0.0002 (sea)	0.005	0.00002	-
Carbon Monoxide	0.03	-	10(9)(8 hr)	10(9) (8 hr)
Chlorate	-	-	-	-
Chloride	-	100.0	-	-
Chlorine, residual	-	0.002	-	-
Chromium	0.05	0.05	0.000002	-
Cobalt	0.0007	1.0	0.0001	-
Copper	0.01	0.005	0.0005	-
Cyanide (as HCN)	0.005	0.005	0.026(0.024)	-
Fluoride (as HF)	-	1.0	-	(0.001)
Hydrogen chloride, HCl (acid), HCl (mist)	-	-	-	-
Hydrogen sulfide	0.002	0.002	0.036 (0.024)	(0.003) (1 hr)
Iron	-	0.3	-	-
Lead	0.010	0.01	0.00036	0.005
Magnesium	0.083	-	0.014	-
Manganese	0.020	0.05	0.012	-
Mercury	0.002	0.002	0.00001	-
Molybdenum	0.070	-	0.012	-

TABLE 7-2. COMPARISON OF MOST STRINGENT STANDARDS CRITERIA WITH MEG CRITERIA FOR AMBIENT BODIES (CONT)

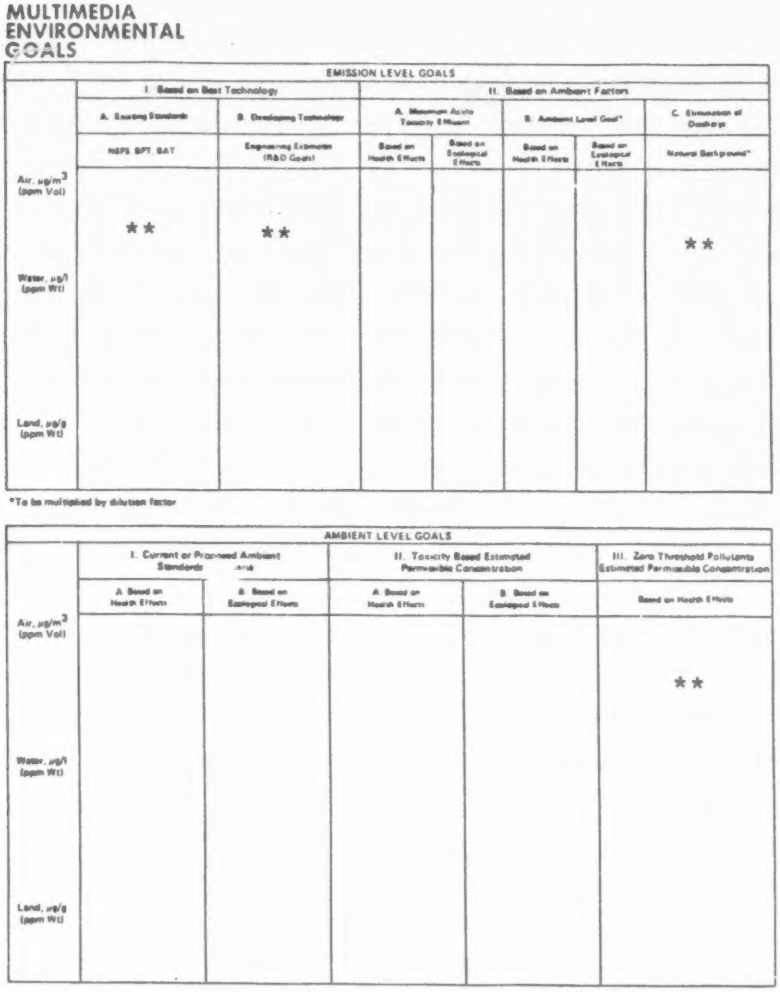
	Maximum Effluent Concentration			
	In Water, mg/l		In Air, mg/m^3*	
Pollutant	MEG MATE**	Most Stringent	MEG MATE**	Most Stringent
Nickel	0.0006	0.025	0.000035	-
Nitrates	-	10	-	-
Nitrates(as N)	-	-	-	-
Nitrogen	-	-	-	-
Nitrogen oxides (as NO)	-	-	-	0.10 (0.05) (annual)
Ozone	-	-	0.16 (0.08) (1 hr)	0.16 (0.08) (1 hr)
Phenols	0.001	0.001	0.045(0.01)	-
Phosphorus	0.0001	0.05	0.00024	-
Polychlorinated biphenyl (PCB), total	0.000001	0.00	0.0012	-
Selenium	0.005	0.005	0.00003	-
Silver	0.005	0.0001	0.000024	-
Sulfate	-	-	-	0.004 (annual)
Sulfides and mercaptans (as S)	0.0138	-	0.0024 (0.001)	-
Sulfur, total reduced	-	-	-	(0.003) (1 hr)
Sulfur dioxide	-	-	-	0.365 (0.14) (24 hr)
Urea	-	-	-	-
Zinc	0.02	0.030	0.0095	-

* Concentrations in parentheses are ppm (v)

**MATE = Minimum Acute Toxicity Effluent. Values shown are the most stringent of the ambient level MEGs for (a) health or ecological effects based on current or proposed ambient standards or criteria or (b) health or ecological effects based on toxicity based estimated permissible concentration

Within the ambient factor category the effluent regulations could be compared with minimum acute toxicity effluent (MATE) values, considering both health and ecological effects, with ambient concentration goals, considering both health and ecological effects or with elimination of discharge values, based on natural background. The latter criterion was generally not available so it was not used. Ambient concentration goals require dilution factors if a comparison is to be made with other effluent criteria such as the most stringent regulations values reported herein. Therefore, the comparison was confined to the MATE criteria for water and air. The more stringent value for this MATE criterion, regardless of whether based on health or ecological effects, was selected.

Ambient level goals were compared wherever data were available. Most stringent receiving water standards in the synopses summary and the most stringent ambient air quality standards from the ambient air section of the report were compared with MEG criteria. As seen in Figure 7-1, the MEG criteria in the ambient level goals chart are based on three main categories of ambient level goals. The ambient standards category and toxicity based category (toxicity based estimated permissible concentrations, or EPCs) take into account both health and ecological effects. Criteria for the third category, zero threshold pollutants, are based only on health effects. Criteria in this category were not generally available for the pollutants involved and thus this category was not used in this analysis. The ambient standards from the MEG charts were presumed to be drawn from a wider body of standards than those synopsized and therefore would be expected to be supported by data on health or ecological effects substantiated as well, if not better, than that data on which the most stringent summary standards are based. Regardless, most of the lower MEG values used for comparison were from category II (toxicity based).



MULTIMEDIA
ENVIRONMENTAL
GOALS

EMISSION LEVEL GOALS							
	I. Based on Best Technology		II. Based on Ambient Factors				
	A. Existing Standards	B. Developing Technology	A. Minimum Acute Toxicity Effluent		B. Ambient Level Goal*		C. Elimination of Discharge
	NSPS, BPT, BAT	Engineering Estimates (R&D Goals)	Based on Health Effects	Based on Ecological Effects	Based on Health Effects	Based on Ecological Effects	Natural Background*
Air, μg/m³ (ppm Vol)	**	**					**
Water, μg/l (ppm Wt)							
Land, μg/g (ppm Wt)							

*To be multiplied by dilution factor

AMBIENT LEVEL GOALS					
	I. Current or Proposed Ambient Standards eria		II. Toxicity Based Estimated Permissible Concentration		III. Zero Threshold Pollutants Estimated Permissible Concentration
	A. Based on Health Effects	B. Based on Ecological Effects	A. Based on Health Effects	B. Based on Ecological Effects	Based on Health Effects
Air, μg/m³ (ppm Vol)					**
Water, μg/l (ppm Wt)					
Land, μg/g (ppm Wt)					

Figure 7-1. Charts for MEGs.

(From "Multimedia Environmental Goals for Environmental Assessment" EPA 600/7-77-136)

**Not used in comparison

The MEG charts now encompass 216 substances with over 650 projected for coverage. Each of these substances is also within one of the 85 categories of substances which EPA has designated for its MEG studies. Work is continuing on MEG charts for many substances now listed in these categories but for which published charts are not yet available. Hazardous substances are added to the designated categories on a continuing basis. The EPA has ranked the 216 substances that are now addressed by MEG charts by two methods to indicate the relative degree of hazard of the substances: a number system from 30 to 4, with the higher numbers more hazardous, and a symbol system using X, XX, and XXX, with XXX substances the most hazardous. Both indicators are shown on the MEG charts. A number of relevant substances, some important and some minor, were found to be addressed by the stringency summaries but not by synopsis in the MEG charts then available.

Several significant limitations briefly mentioned above should be emphasized relative to the comparison and analyses made. This is true as to air contaminants where many of the jurisdictions state effluent regulations in terms of allowable rates (weight or volume per unit of time or per unit of feed entering or Btu input to the facility) as opposed to terms for effluent concentrations as in the MEG criteria. Among others, such is the case with NO_x and SO_2, two important substances for which there are criteria within the National Ambient Air Quality Standards (NAAQS). As seen in TABLE 7-1, the most stringent regulation effluent concentrations are also available for SO_2, but these values are independent of allowable rate regulations which are more common to the jurisdiction studied. Two other NAAQS criteria categories, particulates and non-methane hydrocarbons, encompass a variety of substances and thus are too broad to have comparable substances covered by MEG charts. This is also true of the NAAQS category for photochemical oxidants although some comparison was possible through consideration of ozone concentration criteria only.

Similarly many of the regulations synopsized, especially in the water area, cover physical conditions such as temperature, turbidity, color, or biological criteria such as BOD_5 and total coliform bacteria not addressed by MEG criteria, which cover only elements and substances that can be defined by a chemical formula. Considerable discretion was necessarily exercised in selecting the most stringent criteria for certain contaminants and thus a different result would have been obtained in these cases with the use of slightly different selection guidelines or philosophy. As already stated, and as can readily be seen on the charts herein, there are relevant substances for which one type of criterion, either MEG or most stringent regulation, is available but not both.

Finally, several other MEG factor categories such as elimination of discharge (under emission goals) and zero theshold pollutants estimated permissable concentration (under ambient goals) are still being developed and there generally was not sufficient data of this type available for comparison with the most stringent regulations.

RECOMMENDATIONS FOR PROJECTION OF FUTURE GOALS

Federal, EPA, state, regional, and international environmental standards, present or proposed, have been surveyed and synopsized. From them has been compiled a summary of the most stringent of the air and water regulations. Concentration standards from the stringency analyses have been compared with criteria available for like compounds in the EPA-sponsored study on "Multimedia Environmental Goals for Environmental Assessment" (EPA-600/7-77-136, November 1977, referred to hereinafter as MEG). The stringency study and MEG comparison were limited to the media of air and water as the solid waste synopses contained very few numerical standards for quantitative analysis or comparison. For the same reason the following discussion somewhat emphasizes the air and water media areas; however, much of what is indicated will be indirectly or at least partly applicable to the solid waste category as well.

When the project scope was reduced, planned work to determine projected 2, 5, and 10 year environmental goals, to correspond to the current and proposed standards reported, could not be accomplished. Nevertheless, recommendations can be made as to further work or work methods and approaches to estimate at what level these future goals might be set. The following recommendations are made in an attempt to satisfy this objective. They are not in any order of priority. There would be overlapping between some of the items listed, and it is unlikely that it would be necessary or desirable to pursue all of the approaches.

1. A thorough study of health and ecological effects of applicable substances. Here any number of good sources should be utilized to gather data on criteria including, to name a few, MEG related studies and charts, Threshold Limit Values (TLVs) of various organizations such as ACGIH, NIOSH studies and recommendations, OSHA regulations and reports, U. S. Public Health Service studies, guidelines and standards, NAS/NAE Water Quality Criteria, Chemical Industry Institute of Toxicology reports and reference compilations such as "Industrial Hygiene and Toxicology," by Interscience Publishers. Such a study would also be helpful in better evaluating the current and proposed regulations presented herein, since their controlling basis is generally not known. This is one of, or possibly the highest of, priority recommendations listed here if resources permit such an approach. The MEG comparison made was a preliminary or first step in this direction and could also be expanded on as a more detailed study.

2. Review and analyses of current and proposed applicable regulations of jurisidictions other than those selected. These could include other highly developed countries such as Germany, Japan, France, United Kingdom,and Sweden, other states with either newly discovered or potential coal deposits or known to have very stringent regulations such as California (or Los Angeles County) and other possible international bodies of regulations (the latter not likely to be highly fruitful).

3. Complete process designs of several favored conversion plant configurations with different scenarios for various coal feeds coupled with the use of programmed modelling techniques to determine ambient concentrations of pollutants in the areas outside of facilities and at different altitudes or depths. This would allow complete analyses and comparison of different regulations regardless of their bases or units or their presentation as an equation. Ambient media and effluent regulations could be analyzed equally well.

4. As part of or in addition to Item 1, a thorough study of applicable substances as carcinogens, mutagens, or teratogens and the limitation levels dictated thereby. Studies of the concept of "zero threshold pollutants" as referred to in the MEG report would be recommended here.

5. A "best future technology" approach, based on a study of estimates, forecasts,and reports on developing technology. National and international economic considerations could also be studied and analyses made both with and without them factored in. Projected energy, fuel and transportation availability as possibly affecting such concerns as national security and utility reliability might also be considered.

6. Some of the regulations already included are more indicative of future standards than others. The New Mexico air regulations for gasification plants are probably close to five year levels and possibly somewhat beyond. British Columbia water regulations and International Joint Commission proposed water regulations are probably close to five year levels. Various Water Act mandates are very good indicators for the level of future water standards. "Fishable, swimmable, navigable waters" are mandated by July 1, 1983 and zero discharge of pollutants to navigable waters by 1985. Toxic pollutants in toxic amounts are already prohibited. A closer review of the existing and proposed laws and of the regulations presented herein is highly recommended as an aid in projection of future regulations. A review of the synopses under a different set of guidelines might produce different levels for current most stringent regulations for some substances. Discretion is necessarily involved in the most stringent regulation selection process and philosophy or guidelines used probably resulted in close to the lowest allowable levels possible from such an analysis.

7. Regardless of other approaches or methods pursued, a continuation of the review and updating of applicable Federal and state legislation and regulations as promulgated and published in the Federal Register and other timely periodicals or issuances of the jurisdictions would be necessary. Additional regulations either directly for coal conversion facilities or for other closely related facilities may be promulgated at any time by various jurisdictions and new bases for closely related regulations should also be followed for futher insight. Mandates of new environmental laws as they are passed must be identified and interpreted in the light of possible

effects on regulations or regulation of new parameters. Following new regulations yet to be promulgated under current legislative mandates such as for more ambient air criteria substances and more industries under new source performance standards also will be a necessity.

In air pollution control the effect of the prevention of significant deterioration (P.S.D.) and emission offset regulations, as these control methods mature, on point source regulations for air criteria substances will have to be taken into account. These new regulatory concepts might have a major effect on future regulations of the point source, fugitive emissions, and ambient media type. Mandated changes in state implementation plans due to lack of progress in attaining criteria substance ambient standards must also be reviewed and analyzed for probable effects on other future regulations.

8. Water regulations based on use or specific supply objective requirements (as in Section VI of Canada Federal Water synopses) in various jurisdictions might be used to aid in forecasting receiving water and emission levels regulations. If brought into more general use, this type of regulation might control the setting of receiving water standards just as these generally control point source effluent concentration allowables. Drinking water standards are the best example of water use standards already commonly in effect.

9. Study of relevant substances in the light of elimination of discharge (EOD) type emission level goals, another concept being used in MEG studies. These goals would be the most stringent and are based on the premise that ambient pollu-

tant concentrations should not exceed natural background concentration. Dilution factors are used to put ambient concentrations in terms of effluents. Rural air atmospheres and drinking water and seawater are frequently used in studies to aid in the study indication of natural background concentrations.

SECTION 8

ENVIRONMENTAL DATA ACQUISITION: CONTROL OF LIQUID EFFLUENTS

DEVELOPMENT OF CONVERSION PROCESS EFFLUENT STREAM MODELS

From the study of available information, it was apparent that coal conversion processes could be conveniently divided into three basic types: gasification processes operating at relatively low temperature, gasification processes operating at relatively high temperature and liquefaction processes.

Almost all of the gasification processes that operate at relatively low temperatures (1,035°C or lower) produce phenols, oils, and tars (p/o/t). These processes are best typified by the commercial Lurgi Dry Ash process (143*) since this process produces p/o/t in large quantities, and since more data are available on this process than on any other gasification process. Accordingly, it has been adopted as the base case process for the study of control technology applications. Figure 8-1 illustrates the liquid effluent streams to which control technology must be applied.

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*Reference numbers are Kellogg File Numbers in the project bibliography.

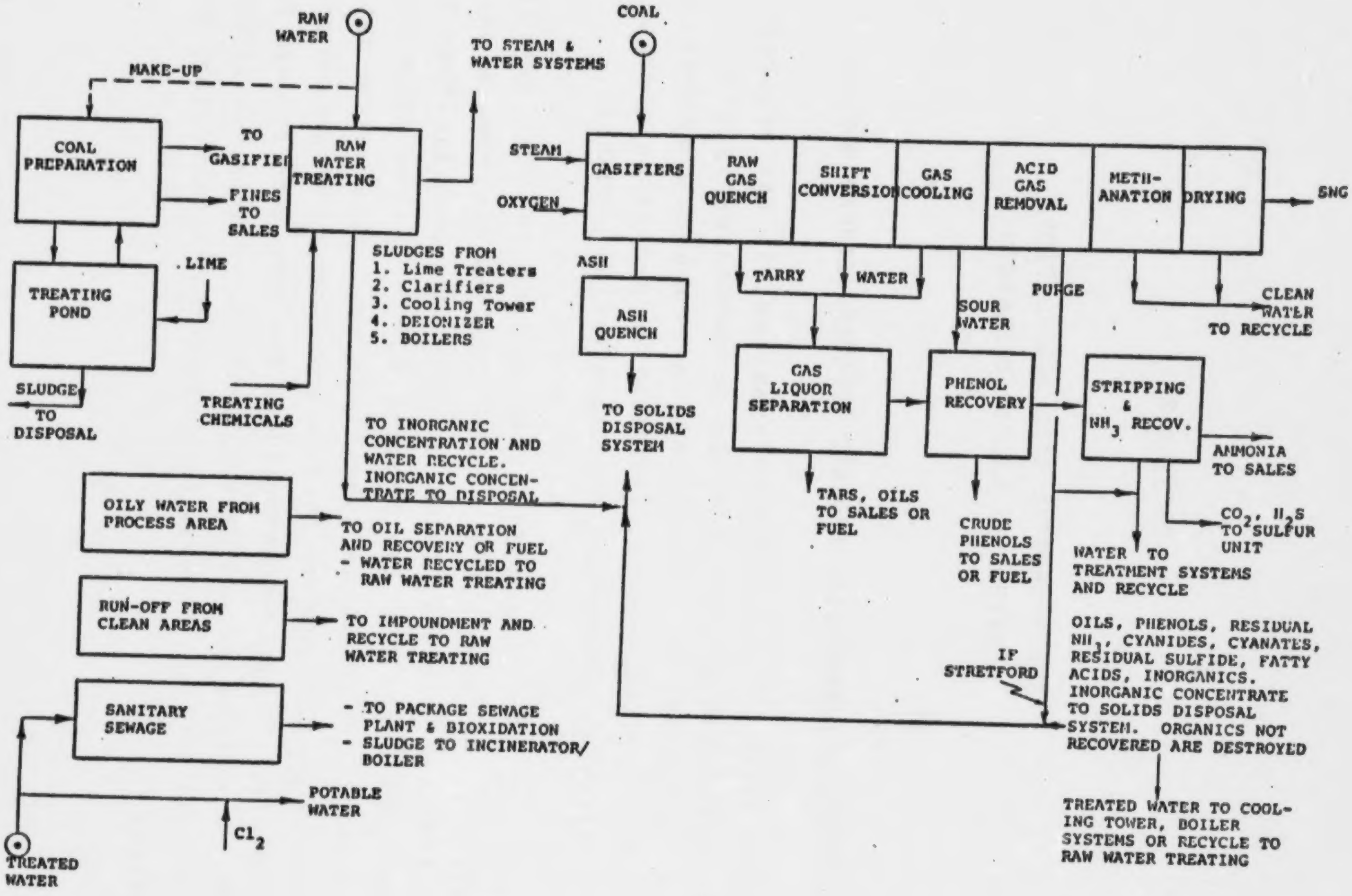

RAW WATER
MAKE-UP
COAL PREPARATION
TO GASIFIER
FINES TO SALES
RAW WATER TREATING
TO STEAM & WATER SYSTEMS
TREATING POND
LIME
SLUDGE TO DISPOSAL
TREATING CHEMICALS
SLUDGES FROM
1. Lime Treaters
2. Clarifiers
3. Cooling Tower
4. DEIONIZER
5. BOILERS
TO INORGANIC CONCENTRATION AND WATER RECYCLE. INORGANIC CONCENTRATE TO DISPOSAL
COAL
STEAM
OXYGEN
GASIFIERS
RAW GAS QUENCH
SHIFT CONVERSION
GAS COOLING
ACID GAS REMOVAL
METH-ANATION
DRYING
SNG
ASH
ASH QUENCH
TO SOLIDS DISPOSAL SYSTEM
TARRY
WATER
SOUR WATER
PURGE
CLEAN WATER TO RECYCLE
GAS LIQUOR SEPARATION
PHENOL RECOVERY
STRIPPING & NH_3 RECOV.
AMMONIA TO SALES
CO_2, H_2S TO SULFUR UNIT
TARS, OILS TO SALES OR FUEL
CRUDE PHENOLS TO SALES OR FUEL
WATER TO TREATMENT SYSTEMS AND RECYCLE
IF STRETFORD
OILS, PHENOLS, RESIDUAL NH_3, CYANIDES, CYANATES, RESIDUAL SULFIDE, FATTY ACIDS, INORGANICS. INORGANIC CONCENTRATE TO SOLIDS DISPOSAL SYSTEM. ORGANICS NOT RECOVERED ARE DESTROYED
TREATED WATER TO COOLING TOWER, BOILER SYSTEMS OR RECYCLE TO RAW WATER TREATING
OILY WATER FROM PROCESS AREA
TO OIL SEPARATION AND RECOVERY OR FUEL
- WATER RECYCLED TO RAW WATER TREATING
RUN-OFF FROM CLEAN AREAS
TO IMPOUNDMENT AND RECYCLE TO RAW WATER TREATING
SANITARY SEWAGE
- TO PACKAGE SEWAGE PLANT & BIOXIDATION
- SLUDGE TO INCINERATOR/BOILER
POTABLE WATER
Cl_2
TREATED WATER

Figure 8-1. Effluent model: Lurgi (p/o/t) gasification

The gasification processes that operate at relatively high temperatures produce little or no p/o/t. These are the entrained flow, slagging ash processes, such as Koppers-Totzek, Bi-Gas and Texaco. We have elected to use the C. F. Braun conceptual design (295) for Bi-Gas, for consistent definition of stream quantities, and the analyses of contaminants from Koppers-Totzek (315), since these are the only available analyses.

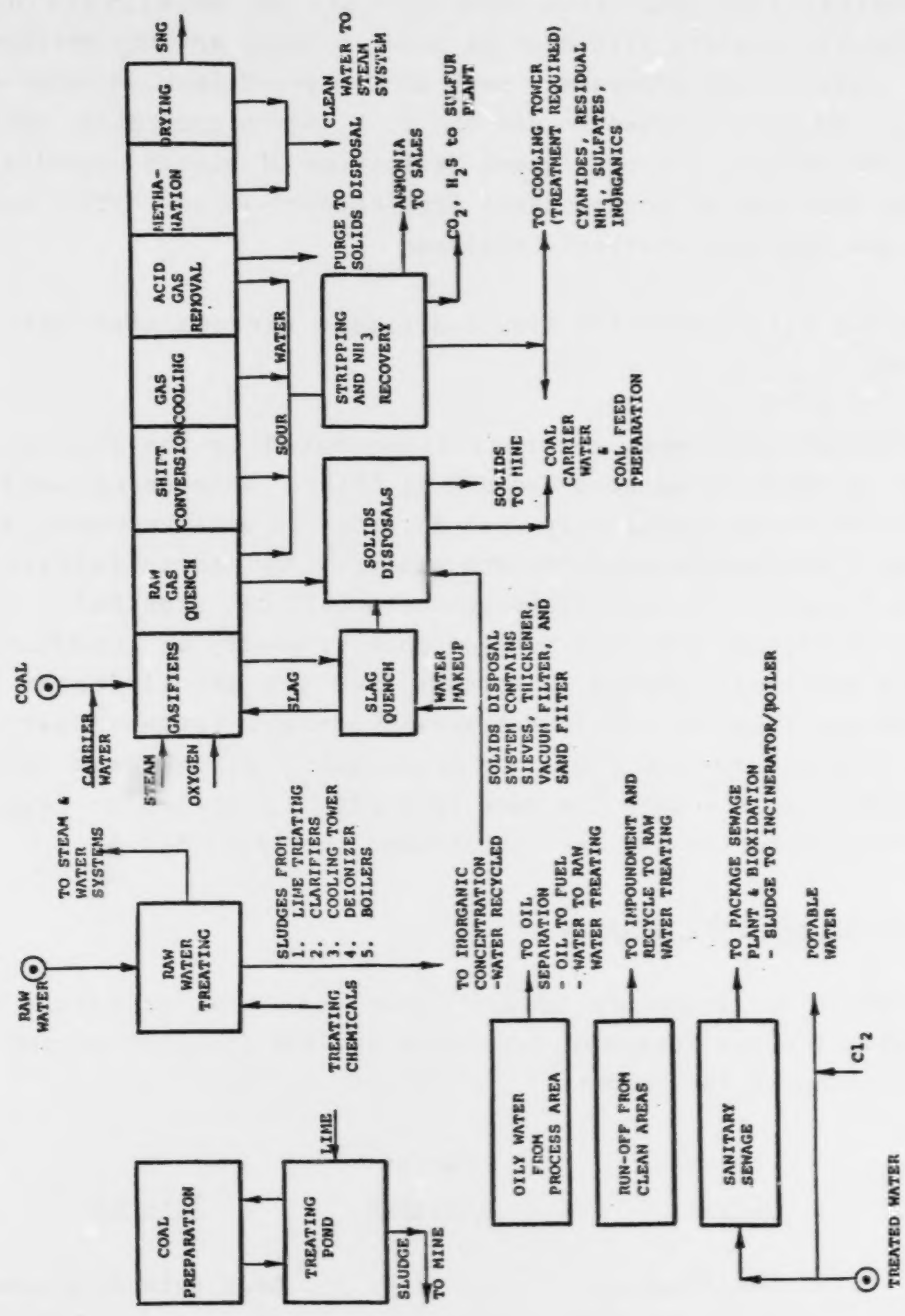
Figure 8-2 illustrates the liquid effluent streams that must be treated.

Liquefaction processes are best illustrated by the conceptual design of Ralph M. Parsons for SRC II (814). This design employs the Bi-Gas Process to gasify coal in order to make hydrogen, both for the liquefaction reactors and also for hydrodesulfurization of the liquefaction distillate products (LPG and naphtha). Fuel gas is manufactured by the Texaco process, gasifying liquefaction residue and coal. Stream quantities from the SRC II design are used in our treating study, but stream contaminant analyses are taken from H-Coal and Koppers-Totzek data, since these latter were more complete than the data from SRC II, Bi-Gas, or Texaco. Effluents to be treated are illustrated in Figure 8-3.

Sour Water Analyses Located

Analyses of contaminated (sour) water have been collected and tabulated for the following processes and are included as part of this section of the report.

Process	Process Type	Number Analyses	Remarks
Lurgi	Gasification. P/o/t Producer.	18	Good data from comcial plants. Four different coal feeds.



COAL
CARRIER WATER
STEAM
OXYGEN
GASIFIERS
RAW GAS QUENCH
SHIFT CONVERSION
GAS COOLING
ACID GAS REMOVAL
METHANATION
DRYING
SNG
CLEAN WATER TO STEAM SYSTEM
PURGE TO SOLIDS DISPOSAL
AMMONIA TO SALES
CO_2, H_2S to SULFUR PLANT
TO COOLING TOWER (TREATMENT REQUIRED)
CYANIDES, RESIDUAL NH_3, SULFATES, INORGANICS
SOUR WATER
STRIPPING AND NH_3 RECOVERY
SOLIDS DISPOSALS
SOLIDS TO MINE
COAL CARRIER WATER & COAL FEED PREPARATION
SLAG
SLAG QUENCH
WATER MAKEUP
SOLIDS DISPOSAL SYSTEM CONTAINS SIEVES, THICKENER, VACUUM FILTER, AND SAND FILTER
RAW WATER
RAW WATER TREATING
TO STEAM & WATER SYSTEMS
TREATING CHEMICALS
SLUDGES FROM
1. LIME TREATING
2. CLARIFIERS
3. COOLING TOWER
4. DEIONIZER
5. BOILERS
TO INORGANIC CONCENTRATION
-WATER RECYCLED
COAL PREPARATION
TREATING POND
LIME
SLUDGE TO MINE
OILY WATER FROM PROCESS AREA
TO OIL SEPARATION
- OIL TO FUEL
- WATER TO RAW WATER TREATING
RUN-OFF FROM CLEAN AREAS
TO IMPOUNDMENT AND RECYCLE TO RAW WATER TREATING
SANITARY SEWAGE
TO PACKAGE SEWAGE PLANT & BIOXIDATION
- SLUDGE TO INCINERATOR/BOILER
POTABLE WATER
Cl_2
TREATED WATER

Figure 8-2. Effluent model: Bi-Gas (no p/o/t) gasification.

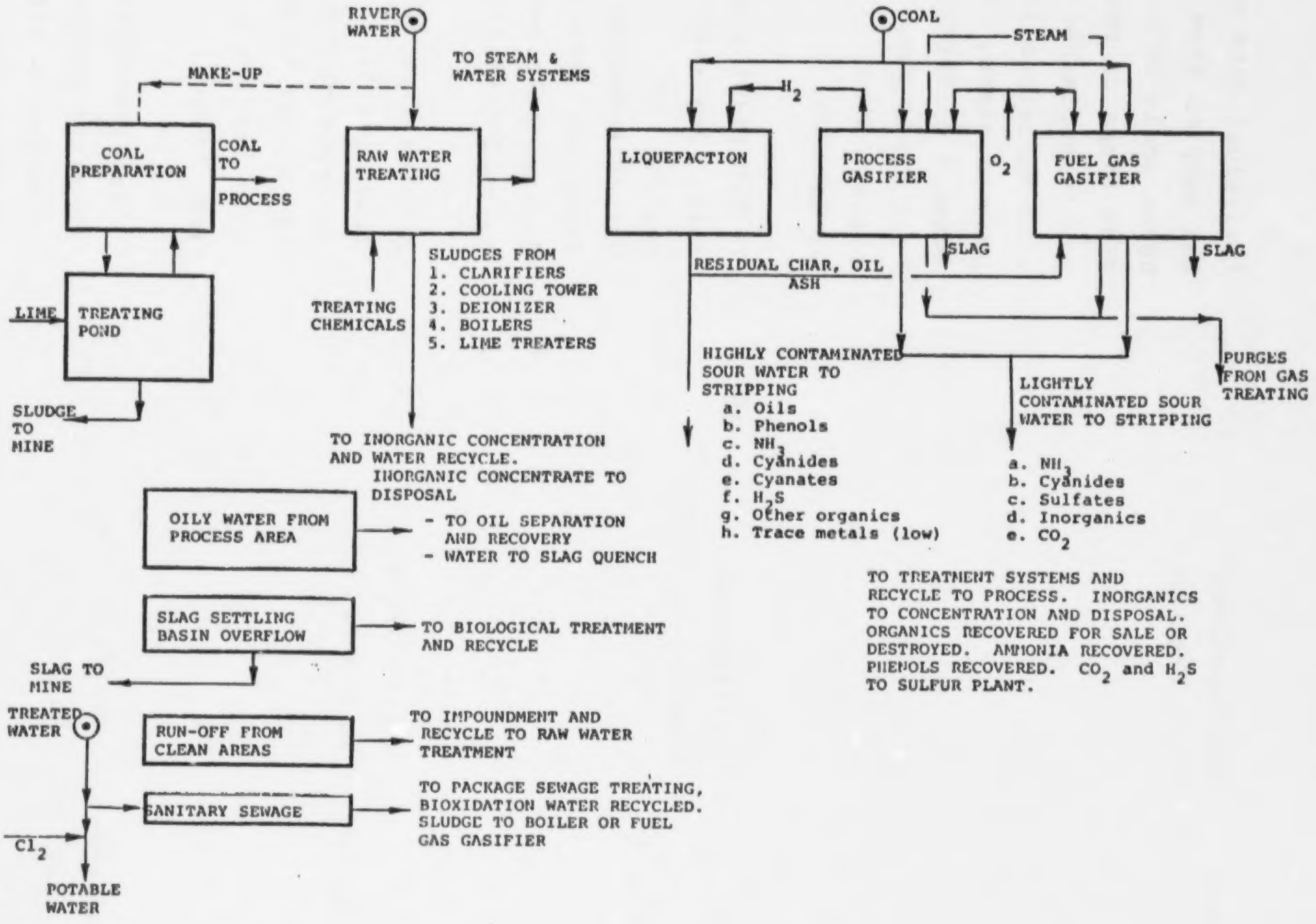

RIVER WATER
MAKE-UP
TO STEAM & WATER SYSTEMS
COAL PREPARATION
COAL TO PROCESS
RAW WATER TREATING
TREATING POND
LIME
SLUDGE TO MINE
TREATING CHEMICALS
SLUDGES FROM
1. CLARIFIERS
2. COOLING TOWER
3. DEIONIZER
4. BOILERS
5. LIME TREATERS
TO INORGANIC CONCENTRATION AND WATER RECYCLE.
INORGANIC CONCENTRATE TO DISPOSAL
OILY WATER FROM PROCESS AREA
- TO OIL SEPARATION AND RECOVERY
- WATER TO SLAG QUENCH
SLAG SETTLING BASIN OVERFLOW
TO BIOLOGICAL TREATMENT AND RECYCLE
SLAG TO MINE
TREATED WATER
RUN-OFF FROM CLEAN AREAS
TO IMPOUNDMENT AND RECYCLE TO RAW WATER TREATMENT
SANITARY SEWAGE
TO PACKAGE SEWAGE TREATING, BIOXIDATION WATER RECYCLED. SLUDGE TO BOILER OR FUEL GAS GASIFIER
Cl_2
POTABLE WATER
COAL
STEAM
H_2
LIQUEFACTION
PROCESS GASIFIER
O_2
FUEL GAS GASIFIER
SLAG
SLAG
RESIDUAL CHAR, OIL ASH
HIGHLY CONTAMINATED SOUR WATER TO STRIPPING
a. Oils
b. Phenols
c. NH_3
d. Cyanides
e. Cyanates
f. H_2S
g. Other organics
h. Trace metals (low)
LIGHTLY CONTAMINATED SOUR WATER TO STRIPPING
a. NH_3
b. Cyanides
c. Sulfates
d. Inorganics
e. CO_2
PURGES FROM GAS TREATING
TO TREATMENT SYSTEMS AND RECYCLE TO PROCESS. INORGANICS TO CONCENTRATION AND DISPOSAL. ORGANICS RECOVERED FOR SALE OR DESTROYED. AMMONIA RECOVERED. PHENOLS RECOVERED. CO_2 and H_2S TO SULFUR PLANT.

Figure 8-3. Effluent model: SRC liquefaction.

Synthane	Gasification. P/o/t Producer.	12	Analytical data good, but samples from bench scale unit. Wide range of results partly explained by wide variation in process steam. At least 6 different coal feeds may explain other variations.
Hygas	Gasification. Phenols, no tars.	16	75 T/D pilot plant data. Analyses reported by Carnegie-Mellon team. Three coal feeds. Contaminants related to coal quantities.
CO_2 Acceptor	Gasification. No p/o/t.	2	Samples from 40 T/D pilot plant. Carnegie-Mellon team and Radian to report more (Radian work was for CONOCO).
Koppers-Totzek	Gasification. No p/o/t.	6	Good commercial data. Organic contaminants low due to high temperature.

U-Gas	Gasification. No p/o/t.	1	Inadequate analysis.
COED	Liquefaction	3	Analyses indicative, but not adequate.
SRC	Liquefaction	8	Pilot plant operation. Samples from full-scale treating system. More analyses are expected to be forthcoming.
SYNTHOIL	Liquefaction	1	More analyses are available, but all are from bench scale operation.
Coal Pile Run-off and Ash Pond Water		13	Coal-fired power plant data.
DOE/MERC	Gasification. P/o/t Producer.	4	Good data from pressurized Wellman-Galusha process.
H-Coal	Liquefaction	1	Good analytical data from bench scale. Treating information available.

GFERC*	Gasification. P/o/t Producer.	1+	Fairly good data. Carnegie-Mellon team providing more data plus treating information. Soon to be published. Process similar to Lurgi Slagging Ash.

The above are indicative of the licensors and developers who have published water analysis data. More is expected, especially from the Carnegie-Mellon project sponsored by DOE/AGA for high Btu gasification processes. Exxon is said to have acquired good data for their Donor Solvent Process and the probability for future publication appears to be good. We expect that more SRC data will be published. Data from proprietary processes and from low Btu gasification processes were not available in time for consideration in this project. Pilot plants near to operation (e.g., Bi-Gas and Battelle) were expected to publish useful data within our project time frame, but publication was delayed.

In TABLE 8-1, entitled "Representative Water Analyses", are shown the best of the water analyses which have been located.

Water Analysis Data Gaps

Among the analyses not located are those from processes now being considered for large scale demonstration plants. These include COGAS, which is an extension of the COED liquefaction work to gasify the char to make hydrogen for the liquefaction, and BCC/Lurgi (Slagging Ash). Both of these were developed by

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* Grand Forks Energy Research Center/DOE.

TABLE B-1. REPRESENTATIVE WATER ANALYSES

Stream	Koppers-Totzek	CO_2 Acceptor		Synthane		HyGas		Lurgi Dry Ash			SRC	H-Coal
	Quench Scrubber	Quench Scrubber		Gasifier Condensate		Gasifier Condensate		Tar Scrubber	Oil Separator		Recycle Process	Liquefac. Condensate
Data Source	Comm.	Pilot Plant		Bench Scale		Pilot Plant		Comm.	Commercial		Pilot Pl.	Bench
Reference (1)	36,315	342		36,230		342		143	143		480	678
	Average	Range	Ave.	Range	Ave.	Range	Ave.	Ave.	Range	Ave.	Ave.	Ave.
BOD mg/1								7200	5200-18200	8800	32500	52700
COD	128	120-300	173	1700-43000	22600			13000	7500-20800	12950	43600	88600
TSS	5084	1290-4825	2705	23-600	151	20-52	35					2
TDS	831	426-1210	748			1352-2168	1782	1880				5300
pH	8.5	7.7-8.7	8.2	7.9-9.3	8.8	7. [illegible].3	7.9	8.8	8.2-9.8	8.6		9.5
Phenols	0.01	<0.001-0.05	0.02	200-6600	3500	14[illegible]-2680	2055	3100	1900-4800	3233	5000	6800
Ammonia	184	1180-1505	1312	2500-11000	8050	2980-5000	4170	13000	13970-17650	15090	7900	14400
Cyanide	12.5	0.02	<0.02	0.1-0.6	0.4	<0.001	<0.001	~ 8	4-14	8	10	(3.7)
Thiocyanate				21-188	101	270-780	510	260	16-193	132		
Chloride	96	42-70	56		500			266	30-210	116		
Nitrate	13.7	0.03-0.07	0.05									
Phosphate	1.21	3.1-11.4	6.1									
Sulfide	0	<0.01-84.1	28.4			120-138	127				10500	29300
Sulfite		10.7-33.3	18.9									
Sulfate	155	56-335	150									
Total S					1400			506	265-1030	804		
Carbonate (2)		400-1998	1200		6000			6000	6550-33930	19250		
Alkalinity (3)	650			HCO_3	11000			11000 (HCO_3 as CO_2)				
Hardness	630	240-2000	880									
Fatty Acid (as Acetic)								560				
Oil	0		0									608
Tar	0		0		Present			21000	500-2200	1228		
Fixed Solids												330
Total N (Kjeldahl)											8300	
Trace Elements Reported	Limited		Yes		Yes		Yes	No		No	Yes	Yes
Total Oxygen Demand												13200
Coals Included (4)	1	2		2,3,4,5,6,10		7		6,8,9,10			11,12	10

(1) Pullman Kellogg reference file numbers
(2) As $CaCO_3$ except Lurgi, which is as CO_2
(3) As $CaCO_3$ except Synthane, which is as HCO_3 and Lurgi which is HCO_3 as CO_2

(4)		
1 - Unknown	5 - W. Ky.	9 - Ill. No. 5
2 - N.D. Lignite	6 - Pittsburg Seam	10 - Ill. No. 6
3 - Wyo. Subbit.	7 - Mont. Lignite	11 - Ky. No. 9
4 - Ill. Lignite	8 - Rosebud	12 - Ky. No. 14

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private funds and only recently are being funded by DOE. Nothing has been published on the emissions or sour water from these processes to date, but it is known that treating systems are being currently designed.

Missing analyses which would be most helpful would include those before and after stripping out CO_2, H_2S, and NH_3. Analysis should always include BOD, COD, all anions and cations, fixed solids, alkalinity, TDS, total nitrogen (Kjeldahl), cyanide, cyanate, TSS, phenols, pH, sulfides, sulfates, TOD, NH_3-N, NO_3-N, hardness, oil, tar, and grease. As can be seen from TABLE 8-1, many of the investigators omit some of these important analyses. Water treating vendors must have all these analyses in order to make reasonable judgments in estimating efficiency of their processes. If they are to actually design and guarantee treating systems they insist on samples for bench scale tests in their own laboratories and, in some cases, pilot plant trials of their equipment on the stream(s) to be treated.

A word should be included on the limitations of some of the analytical methods. For instance, there has always been controversy over the BOD analysis. Certain interfering substances must be removed or suppressed if this analysis is to represent the amount that is biodegradable in an activated sludge system. The five day BOD analysis may not be sufficiently accurate and it is possible that twenty day analyses should always be used. Further, investigation is required to determine whether TOC or TOD analyses should be adopted instead of the BOD analysis.

Cyanide analyses are said to be subject to interference from polysulfides. Phenol analyses do not include all the phenol forms (e.g. para-cresol) and gas-liquid chromatography, which yields more accurate results, is not widely used. Presence of oil, tar, and grease interferes with several analyses and reporting of oils has been spotty. Soluble and insoluble oils exist but have not been reported to any extent.

Our study does not consider treatment of non-aqueous liquids such as the oils and tars from gasification processes and the naphtha, distillate and fuel oil from liquefaction processes. Since all these contain sulfur compounds and phenols, for automotive use these would have to be removed by extraction, hydrodesulfurization, or hydrogenation and for heavy fuel oil usage the sulfur would have to be reduced to 0.3 to 0.5 percent. The treatment procedures are considered to be beyond the scope of this project.

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*Pullman Kellogg Reference number

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LITERATURE SURVEY AND INFORMATION GATHERING

The modus operandi of the project literature survey has been previously described in Section 5 under the headings "Information Procurement, Storage and Retrieval" and "Subjects Monitored".

For water treating, several publications were monitored which apply specifically to this subject, including "Journal of the Water Pollution Control Federation", "Environmental Science and Technology", "Water and Wastes Engineering", and "Industrial Water Engineering". Franklin Institute monitors 4,000 publications for EPA and publishes annually abstracts from these that concern municipal water treating systems.

A number of pertinent articles were found during a survey of the collection of water treating literature compiled by the Pullman Kellogg Chemical Engineering Development Division. Textbooks on coal technology and water treating were acquired or were used at the Pullman Kellogg Research Library. These were supplemented by brochures from water treating process and equipment vendors.

The lag in obtaining reports through NTIS was considerably reduced by personal contacts with Oak Ridge National Library (which publishes DOE reports) or with authors of DOE reports and their DOE project officers. DOE reports are of prime interest, since most of the pilot plant work in progress on coal conversion processes is sponsored, at least partially, by DOE funding. Most of these reports concern progress in operability of the various processes and contain process information with generally small emphasis on analysis or treatment of the wastewater streams.

As mentioned in Section 5, conceptual designs prepared for many of the leading processes were collected for study, especially of the water quantities involved and the treating systems applied to reduce contaminants.

Personal Contacts, Trips, and Meetings by the Water Group

In Section 5 a partial list of contacts and meetings attended was presented. In the list that follows are shown those contacts that yielded water analyses or water treating information.

> Two personal visits with Carnegie-Mellon University representatives regarding their work on an AGA/DOE contract for high Btu processes (Mike Massey, Dick Luthy, Bob Dunlap), supplemented by a number of telephone contacts, correspondence and exchange of reports.
>
> The Pittsburgh Energy Research Center (PERC) at Bruceton, Pa. was visited and the Synthane and Synthoil pilot plants were inspected. Personnel contacted were Ralph Scott, Richard Santore, James Mulvihill, Lloyd Lorenzi, Tom Torkos, and Dr. Sayeed Akhtar. Several reports were acquired at the time and later by telephone contacts with W. P. Haynes and Glenn Johnson.

Through Don Larsen and Robert Culbertson of Dravo Corp. we were permitted to visit the Bi-Gas pilot plant at Homer City, Pa. and talk to Bob Grace of Bituminous Coal Research.

Contacts and correspondence with Russell Perrussel of the Pittsburgh & Midway Coal Mining Co. resulted in our visiting the 50 TPD SRC pilot plant at Fort Lewis, Washington and discussing the water treatment system in operation there.

A number of contacts were made with licensors and vendors through attendance at the Water Pollution Control Federation Conference in Philadelphia, October 2-6, 1977.

Telephone contacts (TC), correspondence (C) and personal meetings (PM) were held with the following:

- Hygas-Lou Anastasia (TC)
- COGAS Development Co. - L.D. Friedman, Ralph Bloom (TC)
- Illinois Coal Gasification Group - Richard McCrary (TC, C)
- H-Coal, (HRI) - John Kumesh (C), Raymond Shutta, S.L. Morris (TC)
- Bi-Gas (Stearns-Roger) E.B. Warnock (TC)
- Argonne National Laboratory - Shern-Yann Chiu (PM)
- DOE-Washington, D.C. - John Nardella (TC, C)
- Arthur G. McKee Co. - Dr. Parsons (TC)
- Davy Powergas, Lakeland, Fla. - I. Marten (TC)
- Chevron Research Co. - J.D. Knapp (TC, C) supplied cost data on wastewater stripping
- George A. Hormel & Co., Minnesota- (TC, C) on cost information on rotating disc contactors

- Calgon Environmental Systems Division, Houston - Janet Thomas (TC, C) on design and cost information on granular carbon clean-up
- Infilco Degremont, Inc., Richmond, Va. - Charles Thornborg, Paul Blue (TC, C) on design and cost information on biological systems, filters, ozonation, etc.
- Zimpro, Inc, Rothschild, Wis. - Claude Ellis (TC, C) supplied design and cost information on PACT (Powdered Activated Carbon Treatment) and Wet Air Oxidation
- Dupont, Wilmington, Del. - F.L. Robertaccio (TC) referred us to Zimpro, Inc. on PACT
- Union Carbide, Houston R. W. Oeben (TC,C) on design and cost information on UNOX and ozonation
- L-A Water Treatment, division of Chromalloy - Robert McCharen, Houston representative and Mike Brunelle of California office (TC, C, PM) supplied design and cost information on ion exchange, reverse osmosis, lime and zeolite softening,and BFW deaerators. Above were for raw water treating but use of wastewater also discussed.
- American Lurgi, Hasbrouck Heights, N.J. Adam Warsh (TC, C) on design and cost information on Phenosolvan process to extract phenols.
- Betz Co. - Don Reed, Houston (TC, C, PM) on necessary treatment and cost to use wastewater as cooling tower make-up.
- U. S. Steel, Pittsburgh, Pa. - R.D. Rice (TC, C) on design and cost of the Phosam-W process.

TARGET POLLUTANT RESIDUALS

In the discussions with water treating licensors and vendors, the summary of most stringent water regulations, as developed by

Pullman Kellogg, for discharge to a receiving body of water was always included. In addition, the generally less stringent requirements for water reuse as cooling tower makeup or as boiler feedwater for high pressure (1,450 psig) steam systems or lower pressure (150-600 psig) steam systems were supplied. Obviously, there are other water uses such as fire water, ash or slag quenching, revegetation of piles of ash or slag or mine refuse, and dust control which do not require high quality water if leaching to ground water is properly prevented by the drainage systems. With the exception of ash/slag quenching, storm water run-off from "clean" (non-process) areas may be utilized for these purposes. This water would be impounded and generally worked back into the coal conversion plant as a substitute for raw water if the quantity is in excess of the uses for revegetation and fire water. In some areas evaporation would account for a substantial part of the work-off of the storm water.

Most licensors, vendors, and process developers of water treating technology take the attitude that total water reuse is the desirable route, since the environmental standards for water discharge are so stringent that such water would probably be better than any raw water makeup supplies available. Nevertheless, licensors,and vendors have been asked to state whether they could meet the most stringent standards for discharge and the costs involved. The general reply has been that this could not be determined without receiving samples for treatment and experimentation. All stated that they are most anxious to receive wastewater samples. Pullman Kellogg was not in a position to supply such samples in the shortened time for this project; however, the Department of Energy is in this position and is urged to proceed with contracts which could result in the necessary experimentation by licensors and vendors by supplying them with wastewater samples or allowing them to bring test equipment directly to the various DOE pilot plants.

TABLE 8-2 shows the target pollutant residuals that were used in discussions with the licensors and vendors. TABLE 8-3 lists additional "most stringent" state environmental requirements, Federal Standards for coal mining and cleaning and U. S. Public Health Drinking Water Standards. It is of interest to note that several state standards are more stringent than drinking water standards, particularly in chlorides and nitrates.

There are non-numerical, unusual, or abstract standards in the laws of some states that are noteworthy because they could be troublesome in obtaining discharge permits. For example:

Toxic Substances	1/20 of 96-hour median tolerance limit (MTL) for persistent toxicants to sensitive indigenous species (Tex., Okla.).
Unnatural sludge and bottom deposits	None
Floating debris	None
Visible oil, grease, or scum	None
Substances harmful or toxic to human, animal, plant, or aquatic life	None
Taste, flavor	Not offensive or unpalatable (Ind.)
Settleable solids	None
Radioactivity (many states)	Gross beta ≤ 100 PCI/l Radium 226 ≤ 1 PCI/l Strontium 90 ≤ 2 PCI/l Dissolved Alpha Emitters ≤ 3 PCI/l

TABLE 8-2. TARGET POLLUTANT RESIDUALS FOR DISCHARGE WATER

(In mg/l unless noted)

ITEM	MOST STRINGENT (STATE)	EFFLUENT STANDARD (Used for Discussion)
Total Dissolved Solids	750 (Indiana)	1,000
Cyanide (Free)	0.005 (Ohio)	0.02
Total Suspended Solids	15-37	37
Phenols	0.001 (Va., Ohio, W. Va.)	0.005
Ammonia	0.15 (Kans.)	2.5-4.0
Biological Oxygen Demand	15-37 (Ill.)	30
Chemical Oxygen Demand	125 (N. Mex)	125
Sulfate	250 (Wyo., Va., Mo.)	600
pH	7.0-8.5 (Wash)	6-9
Alkalinity (as $CaCO_3$)	500 (Va.)	-
Oil, Grease, Tar	None visible	5.0 (Hexane Soluble)
Chlorides	100 (W. Va.)	250
Copper	0.02 (Several)	0.1
Iron	0.3 (Ohio, N. Mex.)	0.3
Chromium	Total 0.01 (Ohio)	Hex. 0.05 Tri 1.0
Zinc	0.075 @ 80 hardness (Ohio)	0.1
Nitrates	10 (Mo., N. Mex.)	10 (as N)
Phosphorus	0.05	0.1
Manganese	0.05 (Ohio, others)	0.1
Arsenic	0.01 (Wyo., W. Va.)	-
Barium	0.5 (W. Va.)	1.0
Beryllium	0.5 (Mo.)	-
Boron	1.0 (several)	1.0
Cadmium	0.005 (Ohio)	0.01
Fluorine	1.0 (Okla., Ky.)	-
Lead	0.04 (Ohio)	0.05
Mercury	0.0005 (Ohio, Ill.)	0.002
Nickel	0.8 (Mo.)	1.0
Selenium	0.005 (Ohio)	0.01
Silver	0.001 (Ohio)	0.05

TABLE 8-3. MISCELLANEOUS WATER STANDARDS

(In mg/l unless noted)

ITEM	MOST STRINGENT (STATE)	FEDERAL (Coal Mining)	1962 Drinking Water (U.S. Public Health)
Alkyl Benzene Sulfonate (ABS)	0.5 (Wyo.)		0.5
Arsenic	0.01 (Wyo., W. Va.)		0.01-0.05
Chloride	100. (W. Va)		250
Copper	0.02 (several)		1.0
Carbon Chloroform Extract (CCE)	0.15 (Va.)		0.2
Fluoride	1.0 (OK., Ky.)		0.6-2.4* Temp. dependent
Iron (Fe)	0.3 (Ohio, N. Mex.)	3.5	0.3
Manganese (Mn)	0.05 (Ohio, others)	2.0	0.05
Nitrate (NO_3)	10 (Mo., N. Mex.)		250*
Cyanide (CN)	0.005 (Ohio)		0.01-
Phenols	0.001 (Va., Ohio)		0.001
Sulfate (SO_4)	250 (Wyo., Va., Mo.)		250
Total Dissolved Solids	750 (Ind.)		500
Zinc (Zn)	as low as 0.075 (Ohio)		0.5
Barium (Ba)	0.5 (W. Va.)		1.0
Cadmium (Cd)	0.005 (Ohio)		0.01
Chromium (Hexavalent)	total Cr 0.001 (Ohio)		0.05
Lead (Pb)	0.04 (Ohio)		0.05
Selenium (Se)	0.005 (Ohio)		0.01
Silver (Ag)	0.001 (Ohio)		0.05
Mercury (Hg)	0.0005 (Ohio, Ill.)		0.002*
Total Suspended Solids		35	
Turbidity - Jackson Units	None - 10 (several)		1*-5 units
Color	None - 15 (Alaska)		15 units
Threshold Odor	None - 3 (Ky., others)		3 units
Coliform Bacteria	10/100 ml (Idaho)		1-4/100 ml
Temperature Rise	2°F		-
Dissolved Oxygen	5-6		-

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*Interim Regulations propose Nitrate (as N) <10. Others note interim (not 1962).

Nitrogen-mg/l (Illinois)	≤ 2.5 April-Oct. ≤ 4.0 other times
Uranium	≤ 5.0 (Va., N. Mex.)
Residual chlorine	≤ 0.5 mg/l
Total dissolved gas	≤ 110% of saturation (several)
Accidental spills (Idah	Contain to prevent pollution (notify department)
Dilution (several states)	Not acceptable as treatment substitute
Methylene blue active substances	≤ 0.5 mg/l (Va.)
Anti-degradation (Va.)	"Water better than standards maintained"
Foaming Agents (Ohio)	≤ 0.5 mg/l

Effluent limitations established for refineries are difficult to relate to coal conversion, since they are expressed in terms of pounds of contaminant per 1,000 barrels of feed stock. However, these are as follows:

BOD	3.1	Ammonia as N	3.0
TSS	2.5	Sulfide	0.017
COD	21.0	Total Chromium	0.049
Oil & Grease	0.9	Hexavalent Cr	0.0032
Phenolics	0.02	pH	6-9

DEVELOPMENT OF THE RECYCLE PHILOSOPHY

Consideration of the number of effluent streams from coal conversion processes, the flow quantities involved and the levels of contaminants in these streams, in view of the stringency of environmental regulations concerning release of effluents to receiving bodies of water, led to several conclusions:

- Proven technology appears to be available for treatment of conversion process effluent streams for reduction of contaminants to levels required by environmental standards for release of effluents to receiving waters.

- The application of treatment technology to the effluent streams from the conversion processes on an individual basis leads to maximum capital investment and operating costs for effluent treatment, maximum raw water usage and maximum problems in disposal of the residual sludges and inorganic salt concentrates from the treatment processes. All of these effects eventually contribute to increasing the manufacturing costs of the conversion process products.

- Problems in meeting environmental regulations, both present and future, can be avoided if there are no effluents from the conversion processes, either by direct discharge or by percolation into subsurface waters.

A simple two-fold philosophy was evolved from these conclusions: Water treatment shall be considered as a means of preparing the individual conversion process effluent streams for recycle as process water within the conversion process battery limits; only the irreducible minimum of water shall be released to receiving bodies of waters.

Among the many advantages associated with the amplification and application of this philosophy the following may be cited:

- Severity of water treatment is minimized, since the treated water must now meet only the standards for process consumption, instead of human (or fauna and flora) consumption.

- Water treatment costs are minimized.

- Raw water usage is minimized.

- Problems associated with disposal of water treatment process residuals are minimized.

- Problems associated with meeting possible future, more stringent, environmental standards are minimized.

There appear to be no disadvantages in adoption of this philosophy. Accordingly, water treatment methods for application to conversion process effluents are not considered individually but rather as parts of a complete treatment scheme with the single objective of minimizing problems, either environmental or economic.

COMMERCIAL WATER TREATMENT METHODS

Commercial water treatment methods are defined as those processes that have been demonstrated in full scale applications. These methods are offered by treatment process licensors and by treatment equipment vendors.

The commercial treatment processes that have been investigated and that are considered to be applicable to the treating requirements in coal conversion processes are:

For Raw Water

(In general these processes deal with inorganic constituents only. Reverse osmosis is an exception.)

Chemical Precipitation (Softening)
Reverse Osmosis
Ion Exchange
Evaporation
Electrodialysis

For Wastewater

(In general these processes deal more with organic constituents than with inorganic constituents, even though both are present.)

Oil Separation
Phenol Extraction
Ammonia Recovery (including stripping)
Chemical Coagulation/Flocculation
Flotation
Biological Oxidation
Filtration
Biological Oxidation Sludge Handling
Carbon Adsorption
Tertiary Treatment (includes disinfection and addition of ozone, chlorine, and hydrogen peroxide)

The individual treatment processes are presented in a general format that includes:

- Description
- Capability, efficiency, and limitations
- Case histories
- Wastes produced
- Cost data (Classical costs only. Later in this section of the report the attempt is made to bring all costs to

the same basis and year so that comparisons on an equal basis are possible.)

- Possible problems
- Possible improvements
- References

It should be understood that there may be an overlap in applications of the treatment processes, in that some of the raw water processes will be used on wastewater when it is recycled for reuse in the process.

An excellent illustration of the effects of variations in raw water quality on treatment methods required and on treatment costs is found in "Water Conservation and Pollution Control in Coal Conversion Processes," D. J. Goldstein and David Yung (Water Purification Associates, Cambridge, Mass.), EPA-600/7-77-065, June 1977 (480*). Selected sites in New Mexico, North Dakota, and Wyoming illustrate the wide variation in raw water quality and supply at probable locations for coal conversion plants. Three types of processes were investigated at each site: liquefaction, gasification to produce SNG, and gasification for power production. Amortized capital costs plus operating costs for total plant water treatment varied from as low as $185 per hour to as high as $835 per hour. Highest cost was for electric power production in North Dakota and lowest for liquefaction in New Mexico. Additional references will be made to this report.

In most of the treatment processes pH is quite important and adjustment of pH with acid, alkali, or CO_2 is often necessary. The importance of pH adjustment and the means of pH control are discussed in "Integrated Schemes for Wastewater Treatment" in this section of the report.

*Kellogg file reference number

Chemical Precipitation

The presence of calcium and magnesium salts in cooling and boiler waters can produce scaling. TABLES 8-4 and 8-5 show limits on the composition of circulating cooling water. Boiler water composition limits are shown in TABLE 8-14 in the description of ion exchange processes.

Softening involves the removal of calcium and magnesium salts by the formation and removal of insoluble precipitates. Lime induces softening as shown in the following reactions (3*):

$$Ca(HCO_3)_2 + Ca(OH)_2 = 2CaCO_3 + 2H_2O$$
$$Mg(HCO_3)_2 + Ca(OH)_2 = Mg(OH)_2 + Ca(HCO_3)_2$$
$$MgSO_4 + Ca(OH)_2 = Mg(OH)_2 + CaSO_4$$

Soda ash or phosphate may be added to remove noncarbonate calcium hardness as shown in the following reactions:

$$CaSO_4 + Na_2CO_3 = CaCO_3 + Na_2SO_4$$
$$3CaSO_4 + 2Na_3PO_4 = Ca_3(PO_4)_2 + 3Na_2SO_4$$

Lime-soda or lime-phosphate softening may be used either as a cold process or as a hot process. The hot process takes advantage of the reduced solubility of $CaCO_3$ at higher temperatures.

Capability, Efficiency, Limitations--

Generally, the concentrations of calcium and magnesium in the effluent of lime or lime-soda softening will be 35 ppm each as $CaCO_3$ equivalent. Iron will be reduced to less than 0.1 ppm and CO_2 will be zero.

* Item 3 in the reference list for this description

TABLE 8-4. OPERATION CONSTRAINTS FOR RECIRCULATING WATER QUALITY*

Characteristic	Constraint	Remarks
Calcium Carbonate	Langelier Saturation Index = 0.0 to 1.0 Ryznar Stability Index = 6.0 to 7.0	Langelier Saturation Index = pH - pH_s Ryznar Stability Index = $2pH_s$ - pH (pH=measured pH pH_s=pH at saturation with $CaCO_3$)
Calcium Phosphate	$(C_{Ca})^3 \times (C_{PO_4})^2 = 0.06$	C_{Ca} = concentration of Ca in mg/l as Ca C_{PO_4} = concentration of PO_4 in mg/l as PO_4
Calcium Sulfate	$(C_{Ca})(C_{SO_4}) = 577,000$	C_{SO_4} = concentration of SO_4 in mg/l as PO_4
Silica	$(C_{SiO_2}) = 150$	C_{SiO_2} in mg/l
Magnesium & Silica	$(C_{Mg}) \times (C_{SiO_2}) = 8,540$	C_{Mg} in mg/l
Suspended Solids	$C_{ss} = 400$	C_{ss} in mg/l
Chloride	$C_{Cl} = 3,000$ (For Stainless Steel only)	C_{Cl} in mg/l as Cl

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*From Item 6 in reference list

TABLE 8-5. CONTROL LIMITS FOR COOLING TOWER CIRCULATING WATER COMPOSITION

	Conventional at low pH*	Suggested at high pH with high concentration and dispersants*
pH	6.5 to 7.5	7.5 to 8.5
Suspended Solids (mg/l)	200 - 400	300 - 400
Ca x CO_3 (as $CaCO_3$)	1,200	6,000**
Carbonates (mg/l)	5	5
Bicarbonates (mg/l)	50 - 150	300 - 400
Silica (mg/l)	150	150 - 200
Mg x SiO_2 (mg/l)	35,000	60,000
Ca x SO_4 (as $CaCO_3$)	1.5×10^6 to 2.5×10^6	2.5×10^6 to 8×10^6
PO_4	---	---
Chlorides***	---	---
NH_3 (mg/l)	---	---

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* From Crits and Glover, "Cooling Tower Blowdown in Cooling Towers", Water & Wastes Eng., April 1975

** More data neded to confirm (footnote from reference above)

*** The chloride limit has been reported as low as 100 mg/l and as high as 3000 mg/l

Other metals may be precipitated as shown in the following list. The presence of other ions may reduce the removal efficiencies reported.

Aluminum - Precipitates as the hydroxide at pH 5.5-7. Halides may cause problems.

Antimony - 90% removal by lime coagulation.

Arsenic - Ferric hydroxide flocculation yields $\leq$0.05 mg/l arsenic levels.

Barium - Ferric and sodium sulfate at pH 6.0 produces precipitation to 0.03-0.27 mg/l levels.

Cadmium - Ferric hydroxide at pH 10 precipitates to levels of 0.1 mg/l.

Chromium - Lime precipitates trivalent chromium at pH 8.5-9.5. Solubility of the trivalent chromium is <0.1 mg/l.

Copper - Lime precipitation at pH 8.5-9.5 produces an effluent of 0.5 mg/l. Cyanide and ammonia may interfere.

Lead - Lime precipitation will remove lead.

Manganese - Significant removal is achieved above pH 9.4 by lime precipitation.

Mercury - Na_2S and NaHS precipitate mercury as HgS. H_2S generation is avoided at pH 10.

Nickel - Lime precipitation removes nickel.

Phosphorus - <0.1 mg/l in the effluent can be produced by two stage lime precipitation at pH 11. Single stage units produce effluents of $\leq$2 mg/l.

Selenium - Removal may be accomplished by reduction to the insoluble elemental form.

Silver - Precipitation with other metal hydroxides in alkaline conditions gives <0.1 mg/l effluents.

Zinc - Lime precipitation reduces effluent concentrations to <1 mg/l.

More data are shown in TABLES 8-6, 8-7, 8-8, 8-9, and 8-10.

Case Histories--

Lime softening has been used for years to soften water for cooling towers and as partial treatment for boiler feed water makeup. Lime addition is also used to adjust pH and induce precipitation of dissolved solids. Ferric salts are also commonly used to aid metal precipitation. Other additives for flocculation, etc. are discussed later in "Coagulation and Floccuation".

Wastes Produced--

The wastes produced by lime softening are the precipitated sludges. Softening sludges may contain calcium carbonate, magnesium carbonates and hydroxides, and calcium phosphates.

When other metals are removed by precipitation, sludges of the various metal hydroxides will be formed. If coagulation aids are used, these will be included in the sludges.

The waste sludges will contain between 0.5 and 5 percent suspended solids which can be concentrated to 10 to 15 percent solids by hydraulic thickeners. Filtration, centrifugation,and evaporation may be used to further dewater sludges. Oily sludges that cannot be de-oiled economically may be incinerated.

Possible Problems--

The presence of ammonia and cyanides inhibits the precipitation of some metals. Since precipitation of the individual metals usually occurs at varying pH's, the optimum pH for metals removal

TABLE 8-6. REMOVAL OF HEAVY METALS BY LIME COAGULATION, SETTLING AND RECARBONATION*

Metal	Concentration Range (mg/l) Influent	Effluent	Removed (%)
Ag	0.24-1.51	0.01-0.02	96-99
As	7.00-8.40	0.20-0.30	96-97
Ba	0.36-1.08	0.04-0.14	87-89
Cd	0.54-5.78	0.01-0.19	95-99
Co	0.42-1.29	0.04-0.09	90-96
Cr^{+6}	0.45-1.40	0.30-1.25	11-33
Cu	0.60-1.47	0.04-0.23	84-93
Hg	3.26-4.45	0.29-0.61	86-91
Mn	1.37-2.26	0.01-0.02	99
Ni	0.75-1.36	0.11-0.20	85
Pb	0.41-1.21	0.04-0.05	90-96
Zn	7.34-9.61	0.12-0.18	97-99

*From: Argaman, Yerachmiel,and Clark L. Weddle, " The Fate of Heavy Metals in Physical-Chemical Treatment Processes," presented at the 75th National AIChE Meeting (June 1973)

TABLE 8-7. REMOVAL OF HEAVY METALS BY LIME COAGULATION AND SETTLING*

Descrip.	Pilot Plant	Pilot Plant		Bench Scale Lab. Tests		Industrial Wastes Plant		Pilot Plant	
Metal	% removal	mg/1	% rem.	mg/1	%rem.	mg/1	% removal	mg/1	% rem.
Ag	-	0.05	97	-	-	-	-	0.013	85
As^{+5}	-	-	-	23.0	54	-	-	-	-
Ba	35	-	-	-	-	-	-	0.43	100
Cd	0	0.016	95	-	-	-	-	-	-
Co	0	-	-	12.0	89	-	-	-	-
Cr^{+6}	87	0.05	9	3.0	0	7630**	99.9	0.17	59
Cu	75	-	-	15.0	79	15700	99.9	0.14	64
Mn	83	-	-	-	-	-	-	0.33	94
Ni	49	-	-	16.0	63	183	99.9	0.015	67
Pb	45	-	-	17.0	97	-	-	0.13	100
Zn	85	-	-	17.0	97	7900	99.9	-	-

* From Item 4 in reference list

**Cr^{+3}

NOTE: Concentrations shown are in the influent wastewater

TABLE 8-8. METALS IN SOLUTION AFTER LIME COAGULATION*

Metal	Calculated Maximum		Experimental				
	pH 11.5	pH 9.6	Run 1	Run 2	Run 3	Run 4	Run 5
Ag	0.007	0.007	0.01	0.02	0.06	-	-
Ba	0.02	0.02	0.04	0.14	0.06	-	-
Cd	0.002	1.42	0.01	0.02	0.19	0.19	0.13
Co	0.002	0.06	0.04	0.05	1.58	-	-
Cu	0.06	0.001	0.04	0.23	0.32	0.31	0.20
Mn	0.003	5.50	0.02	0.001	1.27	-	-
Ni	0.02	0.74	0.11	0.20	1.27	-	-
Pb	1.6×10^{-5}	5×10^{-5}	0.04	0.05	0.25	-	-
Zn	1.60	0.007	0.12	0.18	1.58	-	-

*From Item 4 in reference list
NOTE: All concentrations in mg/l as the metal

TABLE 8-9. REMOVAL OF HEAVY METALS BY FERRIC CHLORIDE COAGULATION AND SETTLING*

	Run 1			Run 2			Run 3			Run 4			Run 5			Run 6		
	In	Out	% Rem.	In	Out	% Rem.	In	Out	% Rem.	In	Out	% Rem.	In	Out	% Rem.	In	Out	% Rem.
Ag	0.01	0.01	-	0.02	0.01	50	0.06	0.02	67	-	-	-	-	-	-	3.30	0.09	97
As	-	-	-	-	-	-	-	-	-	0.30	0.01	97	0.20	0.01	95	-	-	-
Ba	0.04	0.03	25	0.14	0.07	50	0.06	0.09	-	-	-	-	-	-	-	2.30	0.23	90
Cd	0.01	0.01	-	0.02	0.01	50	0.19	0.01	95	0.19	0.04	79	0.13	0.03	77	3.90	0.29	92
Co	0.04	0.017	58	0.05	0.02	60	1.58	0.49	69	-	-	-	-	-	-	3.30	1.40	58
Cr	0.30	0.07	72	1.25	0.63	50	2.53	0.65	78	-	-	-	-	-	-	6.60	3.40	48
Cu	0.04	0.03	25	0.23	0.02	56	0.32	0.29	9	0.31	0.32	-	0.20	0.23	-	1.30	0.17	87
Hg	-	-	-	-	-	-	-	-	-	0.61	0.28	54	0.29	0.03	90	-	-	-
Mn	0.02	0.02	-	0.01	0.01	-	1.27	0.19	85	-	-	-	-	-	-	3.90	2.30	41
Ni	0.11	0.09	18	0.20	0.15	25	1.27	0.66	48	-	-	-	-	-	-	4.60	1.10	76
Pb	0.04	0.04	-	0.05	0.023	54	0.25	0.12	52	-	-	-	-	-	-	2.30	0.23	90
Zn	0.12	0.13	-	0.18	0.04	78	1.58	0.53	66	-	-	-	-	-	-	9.90	1.0	90

*From Item 4 in reference list
NOTE: In and out concentrations in mg/l

TABLE 8-10. METALS IN SOLUTION AFTER $FeCl_3$ COAGULATION. RUN 6.*

Metal	Calculated Max. at pH 8.6	Experimental Run 6
Ag	0.007	0.09
Ba	0.02	0.23
Cd	142	0.29
Co	6.0	1.40
Cu	0.1	0.17
Mn	550	2.30
Ni	74	1.10
Zn	0.07	1.0

*From Item 4 in reference list

NOTE: All concentrations in mg/l as the metal must be determined for each of the chemical precipitation systems.

References--

1. "Development Document for Effluent Limitations Guidelines and New Source Performance Standards for the Steam Electric Power Generating Point Source Category", USEPA 440/1-74 029a, October 1974. *676

2. AWARE, Pretreatment of Industrial Wastewater for Discharge into Municipal Sewers, October 1, 1977. *643

3. Water Quality and Treatment, American Water Works Assn., Inc., Third Edition, 1971.

4. "Water - 1973", AIChE, Argaman and Waddle, No. 136, Vol. 70, 1974.

5. "Water Conservation and Pollution Control in Coal Conversion Processes", Water Purification Assoc., EPA 600/7-77-065, June 1977. *480

6. "Optimal Water Reuse in Recirculating Cooling Water Systems for Steam Electric-Generating Stations", Chen, Y.S., Petrillo, J.L., and Kaylor, F.B., Second National Conference on Complete Water Reuse, Chicago, 1975.

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*Kellogg Reference Number

Reverse Osmosis (1)*

Reverse Osmosis is a process which utilizes a semipermeable membrane and a pressure differential to separate relatively pure water from solutions containing salts, dissolved organics, and colloids. Water is driven through the membrane by pressure leaving behind a concentrated solution of impurities. In order to achieve a separation, the driving force pressure must be greater than the osmotic pressure of the concentrated solution of impurities.

Capability, Limitations, Efficiency--
Operating pressures for reverse osmosis membranes and membrane supports are limited. Spiral wound and hollow fiber membranes cannot withstand more than 600-800 psi. Tubular membranes are limited to about 1,000 psi. Therefore, to allow for sufficient driving force for the separation, the osmotic pressure of the concentrated solution generally cannot exceed 400 psi.

The membranes are limited with regard to pH and temperature, and can be destroyed by high concentrations of chlorine and other oxidizing agents. For example, cellulose acetate membranes are limited to a pH range of 5 to 8 and to a chlorine concentration of 5 mg/l for 15 minutes. Polyamide membranes are limited to a pH range of 4 to 11 and chlorine concentrations of 0.1 to 0.25 mg/l. Both membranes are limited to a maximum operating temperature of 95°F. The solution cannot be concentrated beyond the solubility of the least soluble salt in solution, or fouling and plugging will occur. TABLE 8-11 is a list of operational

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* Numbers refer to the short list of pertinent references at the end of this description.

TABLE 8-11. OPERATIONAL CONSTRAINTS FOR REVERSE OSMOSIS[*]

Characteristics	Constraint	
pH and Hardness	Function of makeup water quality: (1) to control possible chemical deposits (2) to achieve optimal membrane performance	
Suspended solids	<10μ in size	- in feed water
Residual free chlorine	<0.1 mg/l	- in feed water
Total iron and manganese	<0.3 mg/l	- in feed water
Temperature	<95° F	- in feed water
	>35° F	- in feed water
SiO_2	<200 mg/l	- in rejected brine with polyphosphate addition
$CaSO_4$ $\left(C_{Ca} \times \left(C_{SO_4}\right)\right.$	<3,077,000	- in rejected brine with polyphosphate addition
Total dissolved Solids $\left(C_{TDS}\right)$	<20,000-40,000 mg/l	- in rejected brine

[*] From Item 3 in reference list

constraints for reverse osmosis. When used within these limitations, reverse osmosis can be an effective separation process. TABLES 8-12 and 8-13 show some typical ion and group rejections that may be achieved by reverse osmosis.

Pertinent Case Histories--

Reverse osmosis has been used successfully to produce boiler feed water and potable water from brackish water, and as a pretreatment before deionization.

Wastes Produced--

The primary waste produced by reverse osmosis is the concentrated solution of impurities. Another possible waste stream includes the solids from a pretreatment filter. Water in the reject stream can be high (20 to 40 percent of the water treated).

Cost Data--

The following relation is given for determining capital cost for treatment of industrial wastes.

$$C = 1.7 \times 10^6 (Q)^{0.8}$$

C = Capital Cost (1975)
Q = Throughput, MGD (Million Gallons per Day)

However, some sources state that capital cost is a function of impurity level and not flow rate.

Operating Cost--

The following operating cost data is given for municipal waste treatment units:

TABLE 8-12. TYPICAL REJECTIONS BY REVERSE OSMOSIS MEMBRANES*

Species	Rejection (%)
Ca^{2+}	99
Mg^{2+}	99
Na^{+}	94-96
Ni^{2+}	98
Cr^{6+}	95-97
Copper	> 99.5
Cl^{-}	94-95
F^{-}	90
Cyanide	∿90 (pH dependent)
H_3BO_3	30
Sugars	99.9
Formaldehyde	25
Phenol	0-60 (pH dependent)
Benzyl alcohol	0

Rejection =

$$\frac{\text{(concentration at membrane)-(concentration of product water)}}{\text{concentration at membrane}}$$

* From Item 2 in reference list

TABLE 8-13. TYPICAL SOLUTE REJECTION HIGH-SELECTIVITY CELLULOSE ACETATE MEMBRANES*

Solute		Percent Rejection	
	Maximum	Minimum	Average
Calcium Ca^{2+}	99.7	96.3	>99
Magnesium, Mg^{2+}	99.9	93	>99
Sodium, Na^+	97	88	
Potassium, K^+	97	83	
Iron, Fe^{2+} and Fe^{3+}	100	99.9	100
Manganese, Mn^{2+}	100		100
Aluminum, Al^{3+}	99.9	97.3	>99
Chromium, Cr^{6+}+ pH 2.6			92.6
4.2			97.2
7.6			98.6
Ammonia, NH_3	95	77	
Bicarbonate HCO_3^-			80-98
Sulfate, SO_4^{2-}	100	99	>99
Chloride, Cl^-	97	86	
Nitrate, NO_3^-	86	58	
Fluoride, F^-	98	88	
Boron (at pH5)	60	38	
Silica (at pH5)	95	80	
Orthophosphate, PO_4^{3-}	100		>99
Polyphosphate	100		>99
Total dissolved solids (TDS)	99	89	
COD - secondary effluent	97	94	
- sulfite liquor	97.5	94.9	
BOD - secondary effluent	94	81	
- sulfite liquor	92.2	85.8	
Lignin sulfonates	99.4	98.1	
Sucrose			99.9
Proteins			98 to 100
Phenol			Rejection
Acetic acid			Rejection
Glucose			99.5
Color			100
Turbidity			100

* From Item 1 in reference list

Operating Cost [$ per 1000 gallons] (1975)

Size (MGD)	Debt Service	Operating & Maintenance Cost	Total
1	$0.37	$0.36	$0.73
10	0.22	0.21	0.43
100	0.12	0.19	0.31

Care should be taken when applying these numbers, as costs are dependent on the specific application.

Possible Problems--

Irreversible fouling is often the limiting factor on the membrane life, caused by suspended solids in the feed and precipitation of salts. Pretreatment with activated carbon, deep sand filtration, and others may be used. Addition of acid or chelating agents may prevent precipitation in some applications.

The low tolerance of the reverse osmosis membranes to chlorine may represent a very real problem when treating coal conversion plant wastewater.

Possible Improvements--

Determination and development of effective reverse osmosis pretreatment systems would be helpful. Development of pretreatment methods for removal of chlorine or development of chlorine resistant membranes may also be worthwhile. Tests on actual coal conversion streams are required to evaluate the suitability of reverse osmosis for this application.

References

1. "Innovative Technologies for Water Pollution Abatement" by Water Purification Asociates, Prepared for National Commission Water Quality, December 1975. NTIS No. PB-247 390. *612

2. "Physico Chemical Processes for Water Quality Control," by Weber, Walter J., 1972. (Published by Wiley-Interscience) pp. 310-329.

3. "Optimal Water Reuse in Recirculating Cooling Water Systems for Steam Electric-Generating Stations" by Chen, Y.S., Petrillo, J. L., and Kaylor, F. B., 2nd National Conference on Complete Water Reuse, 1975, Chicago.

*Pullman Kellogg Reference File Number

Ion Exchange

Ion Exchange is a process by which ions in solution are reversibly exchanged with ions of an insoluble substance (generally a solid). No significant change in the structure of the solid takes place. The solids range in composition from naturally occurring green sands and bentonite clays to synthetic organic resins and inorganic compounds.

Ion exchange can be used for demineralizing and purifying industrial wastewaters for reuse or discharge, purifying waste liquors for reuse by removal and recovery of metals, and demineralization of raw waters for boiler water makeup. The water quality requirements for boiler feed waters are given in TABLE 8-14.

In ion exchange, a given species of ion is displaced from the exchange material by ions in solution. The exchange material gradually loses its activity and must be regenerated. During regeneration, the original species of ion is replaced in the exchange material and the displaced ions are removed as a concentrated waste solution.

In essence ion exchange purifies the process stream by producing a regenerant waste stream concentrated with the removed ions.

Capability/Efficiency--

Ion exchange may remove cations, anions or both depending on the exchange resins used. Ion exchange may be used in water softening to replace Ca and Mg ions with Na ions. Ion exchange may also be used to produce ultrapure water by replacing cations with hydrogen ions and anions with hydroxyl ions. A water with a conductivity of 0.2 to 0.5 micromhos per centimeter and 0.02 to 0.05 ppm SiO_2 can be produced on a commercial scale.

TABLE 8-14. GUIDELINES FOR WATER QUALITY IN MODERN INDUSTRIAL WATER TUBE BOILERS FOR RELIABLE CONTINUOUS OPERATION*

Drum Pressure (psig)	Boiler Feedwater			Boiler Water		
	Iron (ppm Fe)	Copper (ppm Cu)	Total Hardness (ppm $CaCO_3$)	Silica (ppm SiO_2)	Total Alkalinity*** (ppm $CaCO_3$)	Specific Conductance (micromhos/cm)
0-300	0.100	0.050	0.300	150	700**	7,000
301-450	0.050	0.025	0.300	90	600**	6,000
451-600	0.030	0.020	0.200	40	500**	5,000
601-750	0.025	0.020	0.200	30	400**	4,000
751-900	0.020	0.015	0.100	20	300**	3,000
901-1,000	0.020	0.015	0.050	8	200**	2,000
1,001-1,500	0.010	0.010	0.000	2	0****	150
1,501-2,000	0.010	0.010	0.000	1	0****	100

*From Item 6 in reference list.

**Alkalinity not to exceed 10% of specific conductance.

***Minimum level of OH alkalinity in boilers below 1000 psi must be individually specified with regard to silica solubility and other components of internal treatment.

****Zero in these cases refers to free sodium or potassium hydroxide alkalinity. Some small variable amount of total alkalinity will be present and measurable with the assumed congruent control or volatile treatment employed at these high pressure ranges.

Private Communication: Pullman Kellogg Source

Maximum 60 ppb TDS, 20 ppb Na, 20 ppb SiO_2 in saturated steam.

TABLE 8-15 shows the performance of ion exchange in different applications on the same water. Ion exchange may also be used to remove phenols, ammonia, and heavy metals. Anions are removed based on the following preference:

$SO_4^{=} > CNS^{-} > ClO_4^{-} > I > NO_3^{-} > CrO_4^{=} > Br^{-} > CN^{-}\ HSO_4^{-} > NO_2^{-} > Cl^{-}$
$> HCO_3^{-} > CH_3COO^{-} > OH^{-} > F^{-}$

Ion exchange resins have several limitations. Cation exchange resins cannot withstand temperatures greater than 120-150°C. Anion exchange resins are limited to 30-60°C. Strong oxidizing agents, including nitric acid, chromic acid, chloric acid, and hydrogen peroxide, will degrade the resins. Iron, manganese, and copper in the presence of oxygen may slowly degrade the resins. The resins will become fouled by precipitation or irreversible adsorption of suspended matter, oils, and other dissolved materials.

Rapid exposure to alternating high and low electrolyte concentrations induces osmotic shocks resulting in resin breakage. Removal of a given trace element is not possible without first removing all preferentially exchangable ions.

Case Histories--

- Coke Oven Liquor--Coke oven liquor was treated by activated sludge and then ion exchanged. The ion exchanger feed contained 1200-1300 mg/l ammonia. 96-97% removal was consistently achieved. Ninety percent thiocyanate removal and some color removal also resulted. Gel type

TABLE 8-15. WATER TREATMENT OF THE SAME RAW WATER BY DIFFERENT PROCESSES*

	Raw water	Softening	Partial desalination	Full desalination	
Total hardness, °dh	15	0.1	0.1	0	0
Carbonate hardness, °dH	9	-	-	-	-
Noncarbonate hardness °dH	6	-	-	-	-
Bound HCO_3^-, mg/l	196	196	5-10 mg/lCO_2	5-10 mg/l/CO_2	0
SO_4^-, mg/l	51	51	51	0	0
Cl^-, mg/l	115	115	115	0	0
SiO_2, mg/l	7	7	7	7	traces
Evaporation residue, mg/l	426	426	273	nearly exclusively SiO_2	practically 0

*From Item 4 in reference list

resins cracked but macro-reticular resins did not. Costs (1975) were estimated for a 400,000 gpd flow. Installed cost was $700,000 and included ion exchange, neutralization and coagulation equipment. Operating cost (less amortization) was $1.28/1,000 gal.of which $.94/1,000 gal. was for regeneration.

- Acid Mine Drainage--Potable water was produced from acid mine drainage containing sulfuric acid and ferrous and other ions. The (1975) capital cost for a 500,000 gpd plant was $2,140,000. Operating cost for a 5 million gpd plant was $2.18/1,000 gal. Amortization accounted for 50 to 60% of the operating cost.

Wastes Produced--

The primary waste stream produced in ion exchange is the regenerant waste stream. The quantity of regenerant is a function of resin utilization. Regenerant required can be theoretically calculated from anion and cation analyses. See Item 4 in the reference list. The regenerant chemicals vary from NaCl in water softening to HCl and NaOH in demineralization. The regenerant waste contains, in addition to the desorbed ions from the resin, iron and manganese oxide precipitates, silica,and resin fines.

The other waste stream produced is the periodic spent ion exchange material when change-out is required.

Cost Data--

Costs are dependent on the application. As shown in "Case Histories", capital cost of ion exchange for acid mine drainage was more than that for coke oven liquor.

Possible Problems--

Most problems in ion exchange are due to dirty or fouled beds or improper regeneration. Fouling can be caused by suspended solids in the feed, precipitation of iron, manganese, or copper oxides, dissolved organics,and silica. Presence of these materials in ion exchange feeds could produce problems. Disposal of regeneration wastes could pose a disposal problem.

Use of feed pretreatment units may be required. Evaporators may be needed to handle regenerant wastes.

One source indicates that for solids concentrations over 1,000 mg/l, economics tend to favor reverse osmosis and electrodialysis over ion exchange. Some applications for over 500 mg/l justify reverse osmosis pretreatment before ion exchange to reduce solids feed for ion exchange to 500 mg/l.

Possible Improvements--

Developments to reduce regenerant chemical requirements are desirable. Tests on wastewaters from coal conversion plants are necessary to determine suitability of ion exchange.

References--

1. Innovative Technologies for Water Pollution Abatement. Water Purification Associates, National Commission on Water Quality report. PB-247 390, 1975. *612.

2. Physicochemical Processes for Water Quality Control. Weber, W. J., 1962.

3. Water Quality and Treatment. American Water Works Association, Inc. Third Edition. 1971.

4. Ion Exchangers, Properties and Applications. Dorfner, Konrad. 1972.

5. Manual on Disposal of Refinery Wastes: Volume on Liquid Wastes. American Petroleum Institute. 1968. *808

6. Feedwater Quality in Modern Industrial Boilers - A Consensus of Proper Current Operating Practice. Simon, D. E. Proceedings 36th International Water Conference, Pittsburgh. 1975.

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* Pullman Kellogg reference number

Evaporation

Evaporation may be used to separate water and nonvolatile solids. Evaporation finds application in wastewater treatment, boiler feedwater treatment, and solids dewatering.

Several systems may be used in evaporation. They are 1) solar evaporation ponds, 2) multiple effect evaporation, 3) multistage flash evaporation, 4) vapor compressor evaporation, and 5) oil fluidized evaporation. Oil fluidized evaporation is discussed in another section and will not be covered here.

Solar evaporation involves ponding the water and solids and allowing solar radiation to induce evaporation. The vaporized water may be collected by covering the pond with a clear glass or plastic roof. The evaporated water condenses on the underside of the roof and runs down to a collection system.

Multiple effect evaporation is a fired evaporation system where the heat required for evaporation is reused. See Figure 8-4. The vaporized water is condensed against evaporating water that is at a lower pressure. This water vapor, in turn, is condensed in another effect causing more evaporation. Many effects may be used in a system which optimizes operating costs and capital costs.

Multi-stage flash evaporation (see Figure 8-5) reduces the problem of fouling heat exchange surfaces, a major problem in other fired evaporation systems. Pressurized water is heated to just below the point at which boiling occurs. The hot water is then depressured in stages, resulting in vaporization of the water. The water vapor is then condensed in the feed preheat exchangers.

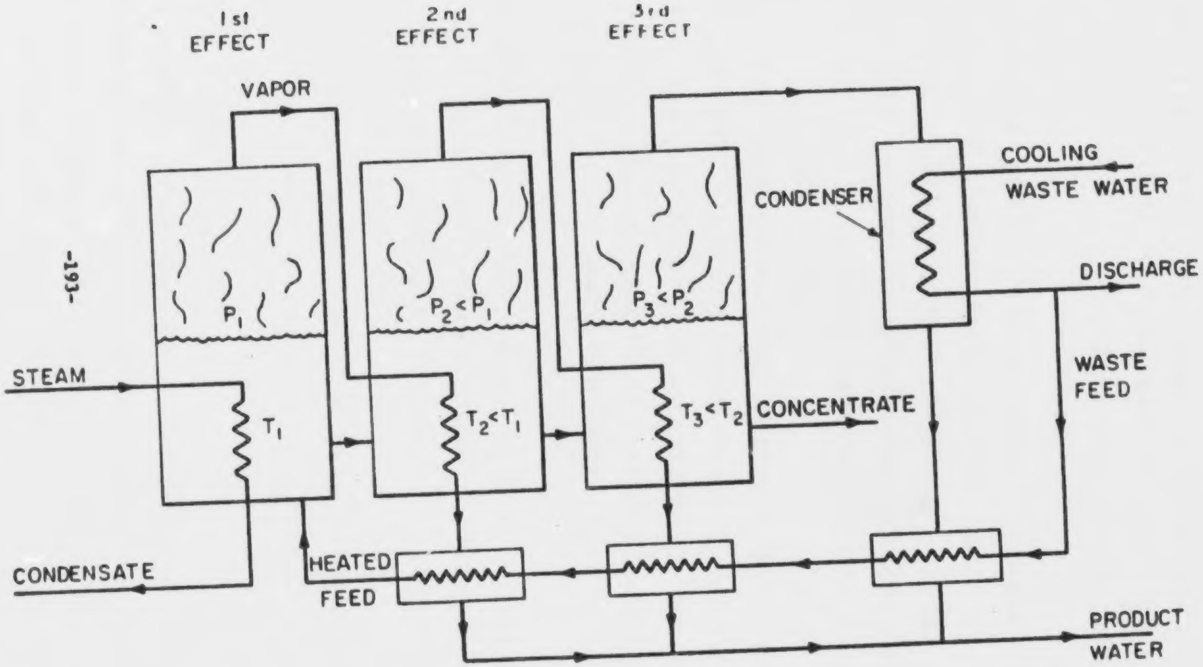
1st EFFECT
2nd EFFECT
3rd EFFECT
VAPOR
P_1
$P_2 < P_1$
$P_3 < P_2$
STEAM
T_1
$T_2 < T_1$
$T_3 < T_2$
CONCENTRATE
CONDENSER
COOLING
WASTE WATER
DISCHARGE
WASTE FEED
CONDENSATE
HEATED FEED
PRODUCT WATER

Figure 8-4. Submerged tube multiple effect evaporator.*

*From Item 1 in reference list

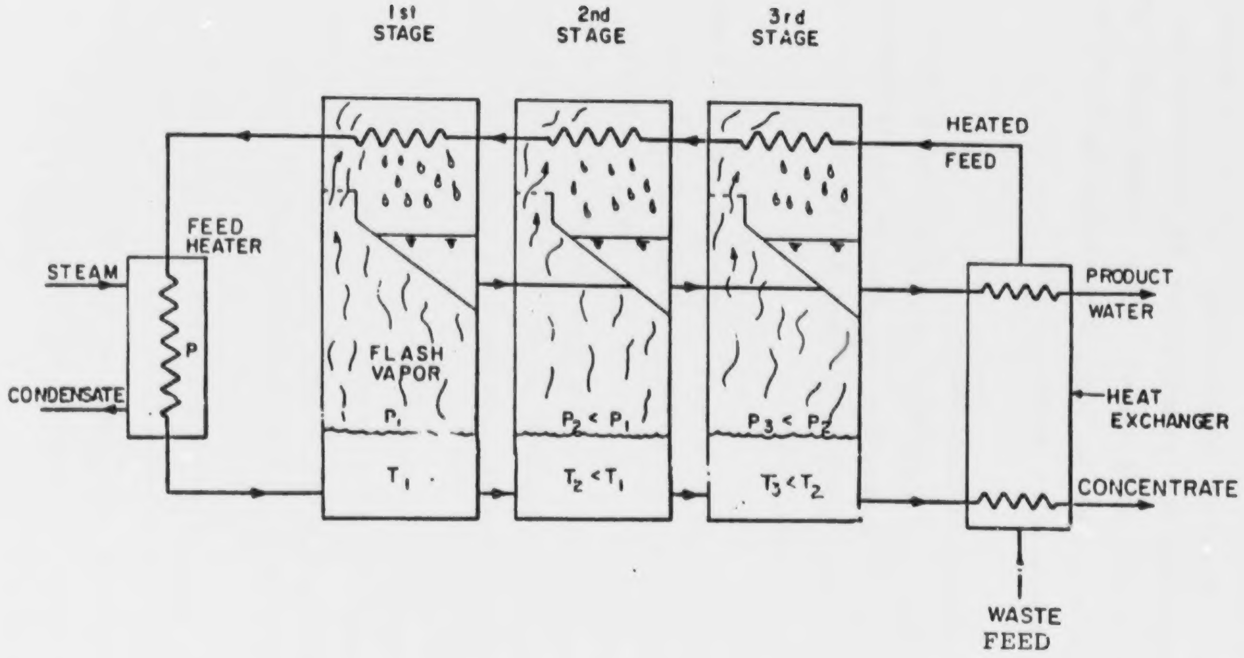
1st STAGE
2nd STAGE
3rd STAGE
HEATED FEED
FEED HEATER
STEAM
CONDENSATE
P
FLASH VAPOR
P_1
$P_2 < P_1$
$P_3 < P_2$
T_1
$T_2 < T_1$
$T_3 < T_2$
PRODUCT WATER
HEAT EXCHANGER
CONCENTRATE
WASTE FEED

Figure 8-5. Multistage flash evaporation.*

*From Item 1 in reference list

Vapor compression evaporates and recovers water by the use of mechanical energy. A vapor compression system is shown in Figure 8-6. The heat required for evaporation is provided by compressing steam formed in the evaporator. The temperature and pressure are increased. This steam is then condensed against water in the evaporators to produce more steam.

Capability/Efficiency/Limitations--
Evaporation systems are capable of handling a wide variety of wastewaters and can concentrate the solids to virtually any level desired. However, evaporation is not to be used on wastewaters containing volatile impurities.

Generally, evaporation systems tend to be extremely energy intensive. The exception is solar evaporation where the energy source is free. Solar evaporation has only limited application when distillate recovery is desired. Large surface area requirements, on the order of 10 to 12 square feet per gallon per day, result in high capital costs.

Energy-wise, vapor compression evaporation is the most efficient of the fired evaporation systems. However, it requires a higher capital investment. Vapor compression evaporation can produce 27 to 40 lbs. distillate/1,000 Btu heat input.

Multiple effect evaporation can produce N lb. distillate/1,000 Btu where N is the number of effects. Energy efficiency must be balanced against capital cost to produce the optimum number of effects. Some rule of thumb figures for number of effects are given in TABLE 8-16.

Multistage flash evaporation generally makes N/3 lbs. distillate/1,000 Btu where N is the number of stages. Multistage flash systems require large recirculation rates to maintain the energy efficiency level. Due to precipitation considerations, multi-

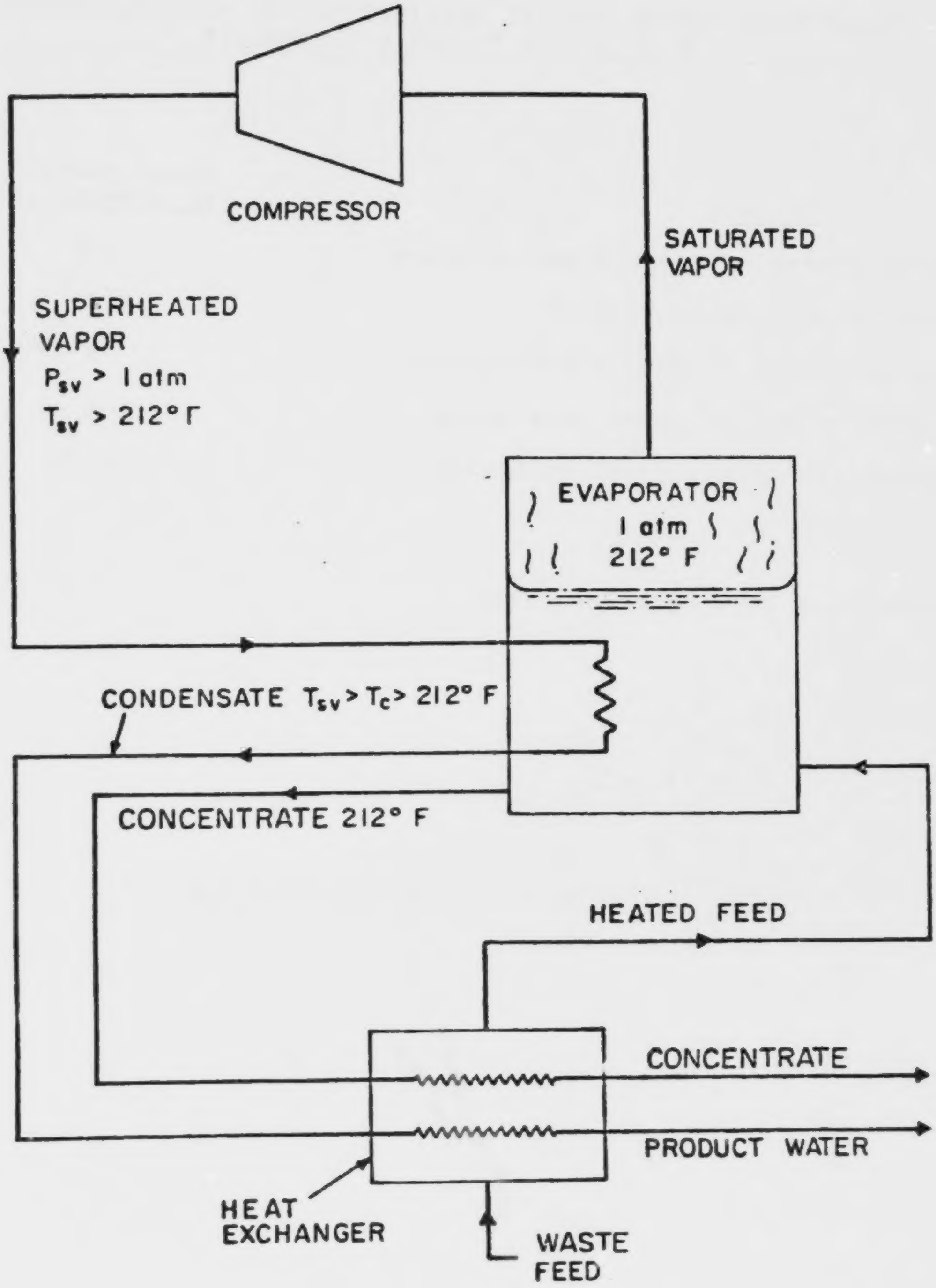
COMPRESSOR
SATURATED VAPOR
SUPERHEATED VAPOR
$P_{sv} > 1$ atm
$T_{sv} > 212°$ F
EVAPORATOR
1 atm
212° F
CONDENSATE $T_{sv} > T_c > 212°$ F
CONCENTRATE 212° F
HEATED FEED
CONCENTRATE
PRODUCT WATER
HEAT EXCHANGER
WASTE FEED

Figure 8-6. Vapor compression evaporator.*
*From Item 1 in reference list

TABLE 8-16. SOME PRESENT APPLICATIONS OF MULTIPLE EFFECT EVAPORATION TO WASTE TREATMENT*

Process	Usual number of effects (N)
Metal finishing recycle and recovery	1-2
Caustic soda concentration	3
Concentration of cane sugar liquors	4
Concentration of paper pulp waste	6
Industrial or municipal salt water	6-20

*From Item 1 in reference list

stage flash evaporation is most efficient on high flow, low TDS applications. The number of stages for sea water desalination for 0.1 to 4 MGD (million gallons per day) is is N = 19 + (Q), where N = number of stages and Q = distillate flow rates in MGD. Sea water desalination plants can currently produce 12 lbs. distillate/ 1,000 Btu.

One important limitation of fired evaporation systems is the fouling of heat exchange surfaces. Multistage flash avoids this problem by using no exchanger surfaces in boiling service. The other methods require the use of antiprecipitant agents, injection of seed solutions which provide alternate precipitation sites, or operation at levels of concentration not exceeding the solubility limit. Therefore heat transfer coefficients of exchanger surfaces have an important effect on efficiency.

The temperature difference driving force is often limited by the solubility limit of the solids in solution.

The use of falling film heat exchangers yields coefficients of 5,000 Btu per hour per square foot per degree Farenheit vs. 300 for conventional tube heat exchangers.

Case Histories--

Vapor compression evaporation has been used to concentrate cooling tower blowdowns. Recovery of purified water is 91 to 98 percent from feed streams containing 1,500 to 10,500 mg/l dissolved solids.

Evaporation has been used to treat sea water, plating bath solution, sugar cane liquors, paper pulp waste, and heavy metal wastewaters.

Wastes Produced--

Evaporation produces pure water and a concentrated solution of dissolved solids. This concentrated stream requires disposal or additional treatment. Precipitation of solids may result in a solids disposal problem. Any precipitation inhibitors added will appear in the concentrate solution. Some manufacturers offer evaporators with crystallizers to facilitate waste handling and disposal.

Costs--

For streams of 0.5 to 3.0 MGD, capital costs are around $2 per gallon per day. Operating costs are shown in Figure 8-7 for multistage flash units.

Vapor compression evaporation units have lower operating costs and higher capital costs. Solar evaporation (when used to produce purified water) has capital costs of $10 to $25 per gallon per day if provided with covers and collection means for the water (1).

Possible Problems--

The primary problem with fired evaporation systems is the high energy requirement. Evaporation does not eliminate a waste stream but merely concentrates it. The concentrated stream may pose a disposal problem. Other problems are fouling and scaling, especially when high recoveries are desired.

References--

(All unreferenced material was taken from Reference 1.)

1. "Innovative Technologies for Water Pollution Abatement," Water Purification Associates. NTIS PB-247 390, Dec. 1975. *612

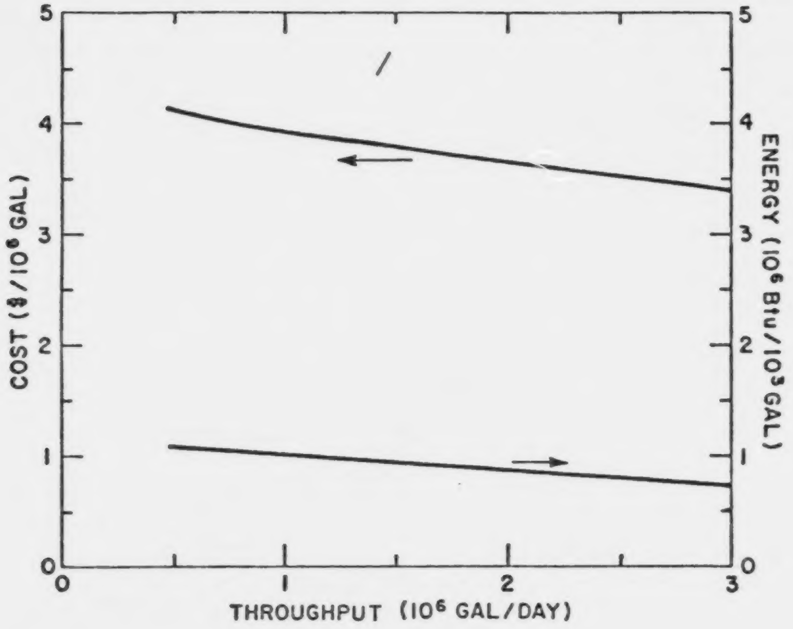

Figure 8-7. Cost and energy for multistage flash evaporation.*

*From Item 2 in reference list

2. "Water Conservation and Pollution Control in Coal Conversion Processes," Water Purification Associates, EPA 600/7-77-065, June 1977. *480

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*Kellogg Reference Numbers

Electrodialysis

Electrodialysis utilizes a DC electric field and semipermeable membranes to remove ions from water. An electrodialysis unit is shown in Figure 8-8. A series of parallel membranes separated by channels is used. The parallel membranes alternate in type between cation-permeable membranes and anion-permeable membranes. The electric field causes the ions to migrate toward their oppositely charged electrodes. The ions pass through the adjacent membrane into the next channel where a concentrated brine solution of the ionic impurities is formed. Purified product water is therefore produced in every other channel This series of channels and membranes is referred to as a "stack." A stack may have as many as 100 to 750 channels between a single pair of electrodes. Stacks may be used in series to produce better effluents.

Capability/Efficiency/Limitations--

Electrodialysis is capable of removing only charged ionic particles. Unlike reverse osmosis, little or no effect on uncharged dissolved particles and suspended matter is observable. The process is very energy efficient with energy costs proportional to impurity of the feed. The primary economic consideration is membrane cost and maintenance. Membrane fouling, particularly the anion-permeable membrane, is a problem. Since the membranes are ion exchange resins made in sheet form, removal of particular contaminants by pretreatment may be required as in ion exchange.

For desalting brackish waters, it has been determined that treatment to not less than 500 mg/l of salts in the product water is economically optimal. Additional stages should be used if lower salt contents are desired. The power consumption required

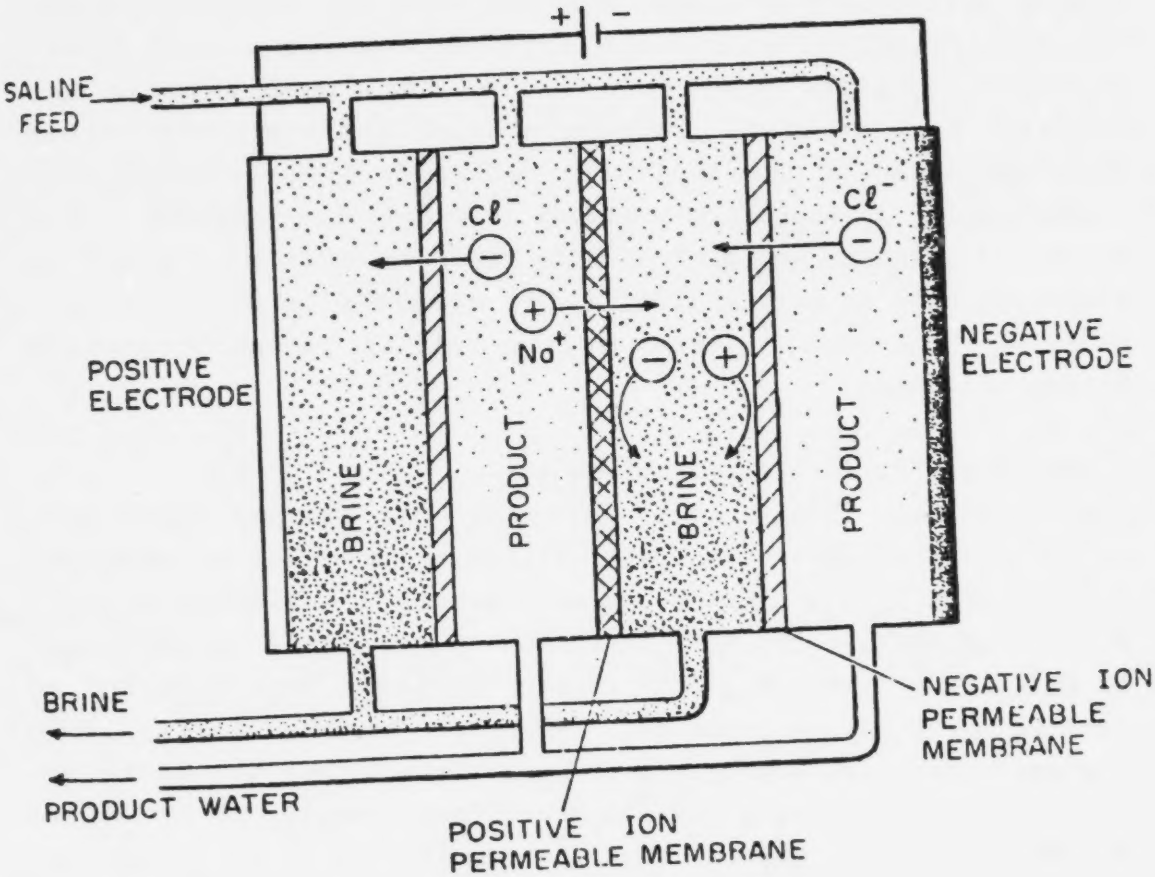
+
−
SALINE FEED
Cl⁻
Cl⁻
Na⁺
POSITIVE ELECTRODE
NEGATIVE ELECTRODE
BRINE
PRODUCT
BRINE
PRODUCT
BRINE
PRODUCT WATER
NEGATIVE ION PERMEABLE MEMBRANE
POSITIVE ION PERMEABLE MEMBRANE

Figure 8-8. Electrodialysis.*

*From Item 1 in reference list

for desalting brackish water to 500 mg/l is shown in the table below. Pumping power required is about 2 Kwh/ 1,000 gal.

The concentration of the brine solution is limited by the possibility of precipitation onto the membranes. Chemical agents may be added to help prevent precipitation.

Calculated Power Consumption for Desalting Brackish
Waters by Electrodialysis
(excluding internal pumping power)

Feed Concentration (mg/l)	Power Consumption (Kwh/ 1,000 gal)
4,900	19
4,000	14
3,000	10
2,000	5

Case Histories--

Electrodialysis has been widely used to desalinate brackish waters. Electrodialysis has also been used to treat metal finishing wastes to 500 mg/l TDS. Ion exchange was used for polishing.

Wastes Produced--

The primary waste stream produced is the concentrated brine stream. Chemical cleaning of the membranes may produce wastes of low pH.

Cost Data--

Power costs for electrodialysis are not high. Power requirements for treating brackish water were given previously. Capital costs for desalting brackish water to 500 mg/l TDS are given in Figure

8-9. The operating cost for a 1.5 MGD plant is about $0.29/1,000 gal. (1975).

The capital cost for a 7000 GPD unit treating waste pickle liquor from a steel plant was $245,000 with an operating cost of 4.5¢/gal. (1975).

Possible Problems--

Membrane fouling is a problem experienced by electrodialysis. Pretreatment of the feed wastewater and membrane treatment as in ion exchange may be required.

The primary costs are for the membranes and their maintenance. Therefore, treatment to lower than 500 mg/l TDS may become prohibitively expensive due to increased membrane requirements.

Possible Improvements--

At low salt concentrations, the economics of the process may be improved by improving flow conditions through the electrodialysis unit. Development of membranes resistant to fouling will aid in reducing operating costs.

Reference--

"Innovative Technologies for Water Pollution Abatement," Water Purification Associates, NTIS PB-247 390, Dec. 1975. (Kellogg Reference Number 612)

1 = One Stage, approximately 50% demineralization

2 = Two Stages, approximately 75% demineralization

3 = Three Stages, approximately 87.5% demineralization

4 = Four Stages, approximately 93.8% demineralization

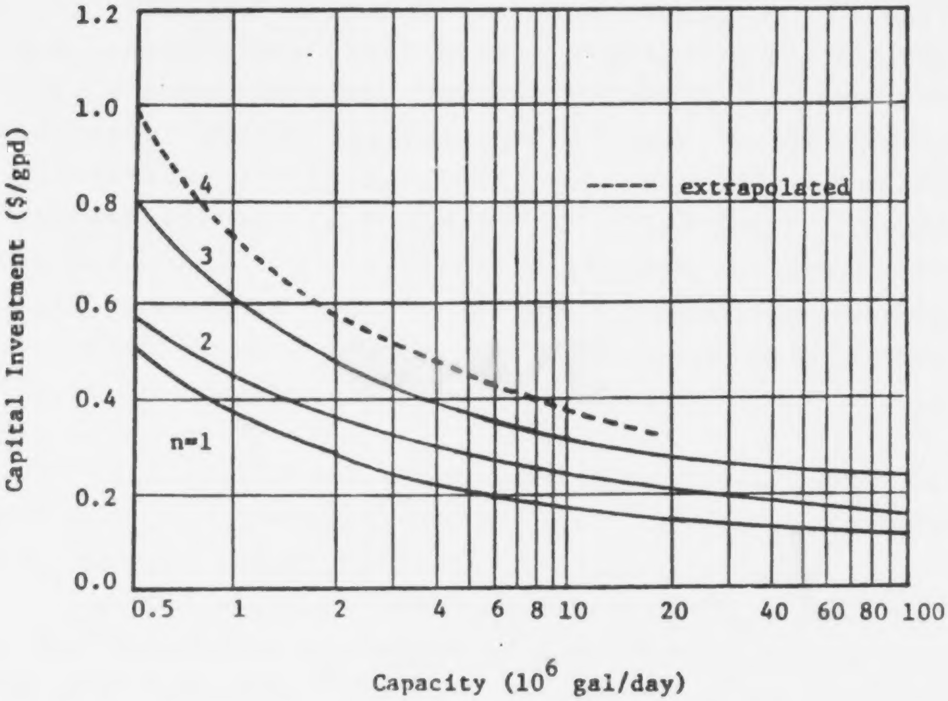

Figure 8-9. Capital investment for electrodialysis as a function of capacity and of number of stages.*
*From Item 1 in reference list

Oil-Water Separation

Oily materials may be free, emulsified, or dissolved in the wastewater but only free and emulsified oils can be separated and removed by physical-chemical methods.

API Separators--

The American Petroleum Institute (API) sponsored studies and published design methods (1*) for oil-water separators based on the rise-rate (Stokes velocity) of oil droplets of 150 micrometers diameter. The API reference should be studied for specific design details and their derivation. In general, application of methods yields long, rectangular, multichannel separators. A Stanford Research Institute report (2) shows the effect of oil specific gravity and system temperature on a 4 million gallon per day (MGD) design API separator. Data are shown in TABLE 8-17. In TABLE 8-17 all of the designs for oils with specific gravity greater than 0.9 (room temperature API gravity of about 25°) are 100 or more feet long. Figure 8-10 is a drawing of an API separator showing some details of construction. The SRI report describes API Separator construction:

> "The walls of the chambers are 7 feet high for a 5 foot water depth. The preferred material of construction is reinforced concrete, although steel separators have been built. The separator chambers include a covered 5 foot x 22 foot preseparator, a floating oil skimmer section 12 feet x 25 feet, a forebay 12 feet x 30 feet and twin settlers 15 feet x 130 feet."

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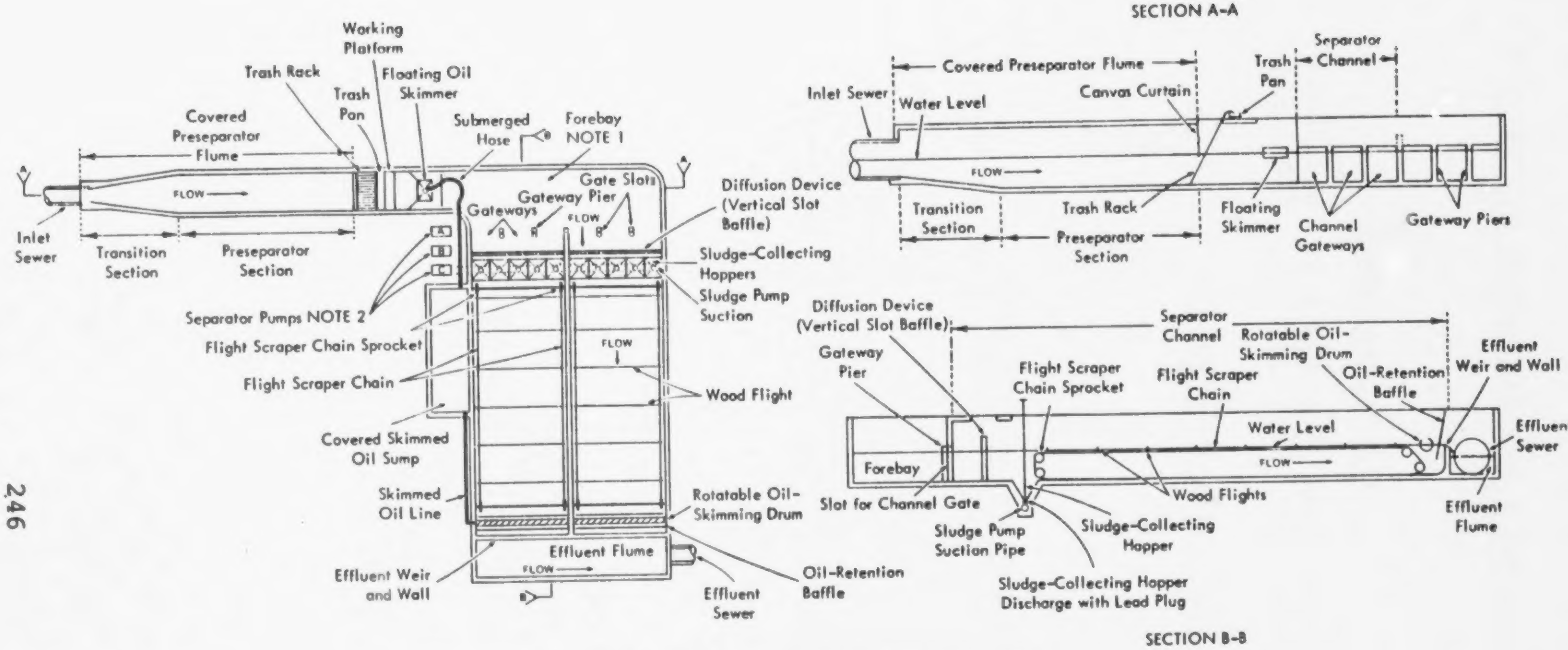
*Item 1 in reference list

TABLE 8-17. API SEPARATORS FOR 4 MGD WASTEWATER DESIGN FLOW

Oil in Feed		Rise Rate (Stokes Velocity), fpm	API Separators				Residence Time, min
Specific Gravity	API Gravity		Number	Dimensions,ft W	D	L	
		68°F, 0.01 Poise Pure Water					
0.9659	15	0.085	2	20	7.5	180	146
0.9340	20	0.16	2	15	5	130	53
0.9042	25	0.23	2	13	5	100	35
0.8762	30	0.30	2	13	5	75	26
0.8498	35	0.36	2	13	5	60	21
		40°F, 0.0154 Poise Pure Water					
0.9042	25	0.14	2	17	5.5	130	66


SECTION A-A
Working Platform
Trash Rack
Trash Pan
Floating Oil Skimmer
Covered Preseparator Flume
Submerged Hose
Forebay NOTE 1
Gate Slots
Gateway Pier
Gateways
FLOW
Inlet Sewer
Transition Section
Preseparator Section
Diffusion Device (Vertical Slot Baffle)
Sludge-Collecting Hoppers
Sludge Pump Suction
A
B
C
Separator Pumps NOTE 2
Flight Scraper Chain Sprocket
Flight Scraper Chain
Wood Flight
Covered Skimmed Oil Sump
Skimmed Oil Line
Rotatable Oil-Skimming Drum
Effluent Flume
Effluent Weir and Wall
Effluent Sewer
Oil-Retention Baffle
Covered Preseparator Flume
Inlet Sewer
Water Level
Canvas Curtain
Trash Pan
Separator Channel
Transition Section
Trash Rack
Preseparator Section
Floating Skimmer
Channel Gateways
Gateway Piers
Diffusion Device (Vertical Slot Baffle)
Separator Channel
Rotatable Oil-Skimming Drum
Effluent Weir and Wall
Gateway Pier
Flight Scraper Chain Sprocket
Flight Scraper Chain
Oil-Retention Baffle
Water Level
Effluent Sewer
Forebay
Slot for Channel Gate
Wood Flights
Effluent Flume
Sludge Pump Suction Pipe
Sludge-Collecting Hopper
Sludge-Collecting Hopper Discharge with Lead Plug
SECTION B-B

NOTE 1: SLUDGE REMOVAL FROM FOREBAY BY CLAM SHELL OR VACUUM TRUCK

NOTE 2: SEPARATOR PUMPS ARE ALLOCATED FOR SPECIFIC SERVICE, AS FOLLOWS, BUT ARE MANIFOLDED TO PERMIT SPARING AND TO PROVIDE FLEXIBILITY OF OPERATIONS

PUMP A: TRANSFER OF SETTLED WATER FROM COVERED SKIMMED-OIL SUMP TO INLET OF PRESEPARATOR FLUME

PUMP B: TRANSFER OF SEPARATED OIL FROM COVERED SKIMMED-OIL SUMP TO SLOP HANDLING FACILITIES

PUMP C: TRANSFER OF SLUDGE FROM SEPARATOR TO SLUDGE HANDLING FACILITIES

Figure 8-10. API oil-water separator.*

*From Item 2 in reference list

Figure 8-11 shows the effects of design flow rate and actual flow rate on operating costs.

Tubular Separators--

In ideal flow, settling is related to flow per unit surface area and is independent of liquid depth. Thus, closely spaced surfaces decrease the separator length needed for collection of a given size particle (2). Tube settlers are a successful design based on this theory. Figure 8-12 shows a cutaway of a steeply inclined tube settler that might be used in wastewater treatment.

CPI Separators--

The Corrugated Plate Interceptor (CPI) separator was developed for the removal of oil as well as solids and has performed well in this service. Corrugated fiberglass reinforced plastic (FRP) plates are stacked with about 0.8-inch spacing and slanted at 45° (2) as shown in Figure 8-13. For oil separation the main flow is downward with coalescing oil droplets gathering on the upper surfaces and rising to flow countercurrently and exit near the inlet of the CPI.

Figure 8-14 shows the effect of design flow rate and actual flowrate on operating costs. In addition to a gross reduction in required land area, CPI separators are estimated to require 20% less capital cost than comparable API separators (2).

Efficiency--

Based on the theory that settling efficiency is related to flow-per-unit surface area, Thompson (3) presented the graph of effluent oil concentration versus the reciprocal of flow per unit area that is shown in Figure 8-15. In the figure, operating data with influent waters containing more than 1,000 mg/l oil were plotted as effluent oil concentration versus the area-to-flow

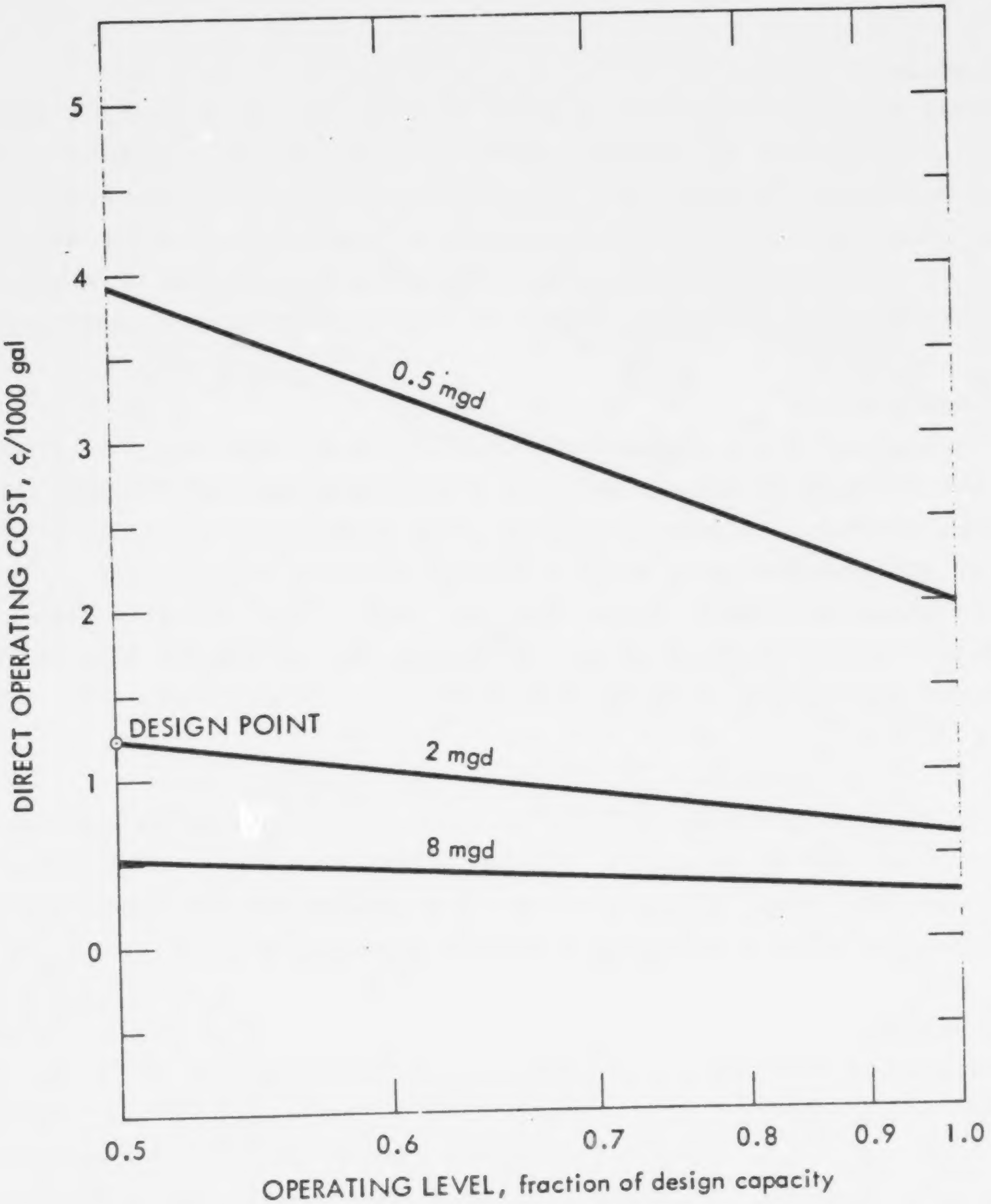


Figure 8-11. API Separator: effect of operating level and plant capacity on operating cost.*

*From Item 2 in reference list

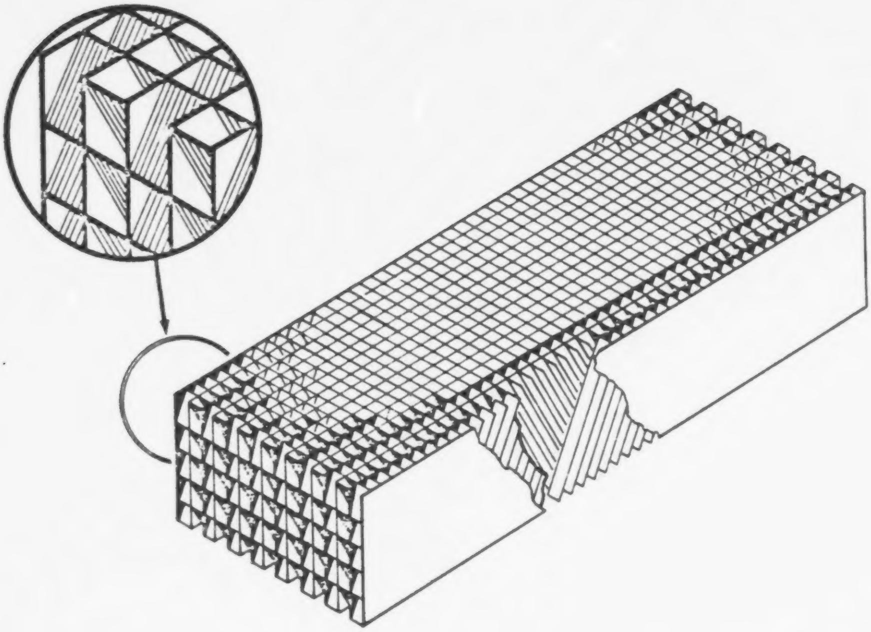


Figure 8-12. Module of steeply inclined tubes.*

*From Item 2 in reference list

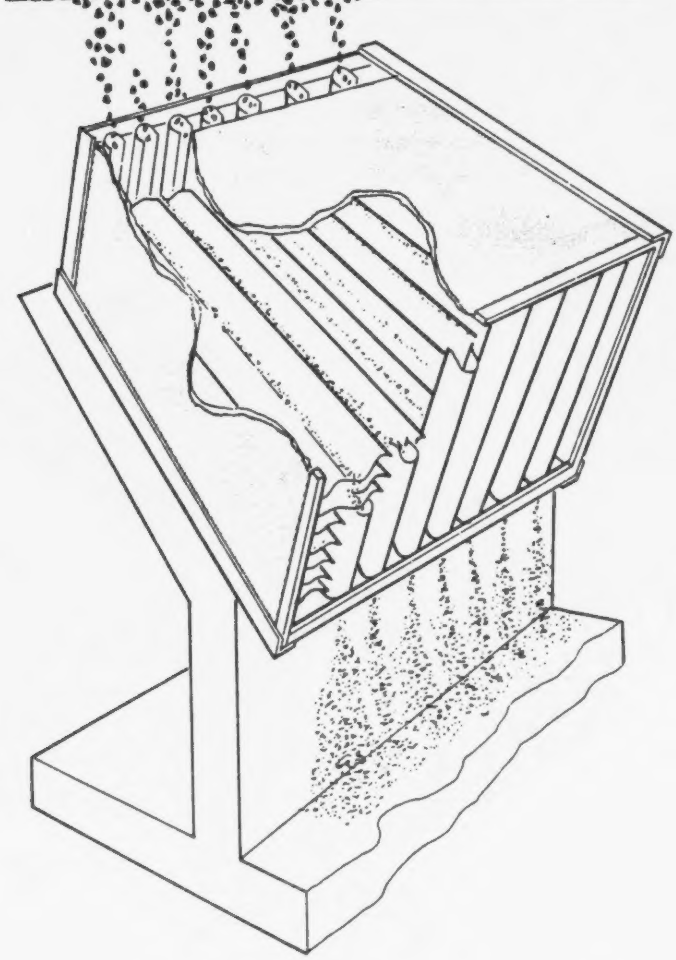


Figure 8-13. Corrugated plate interceptor.*

*From Item 2 in reference list

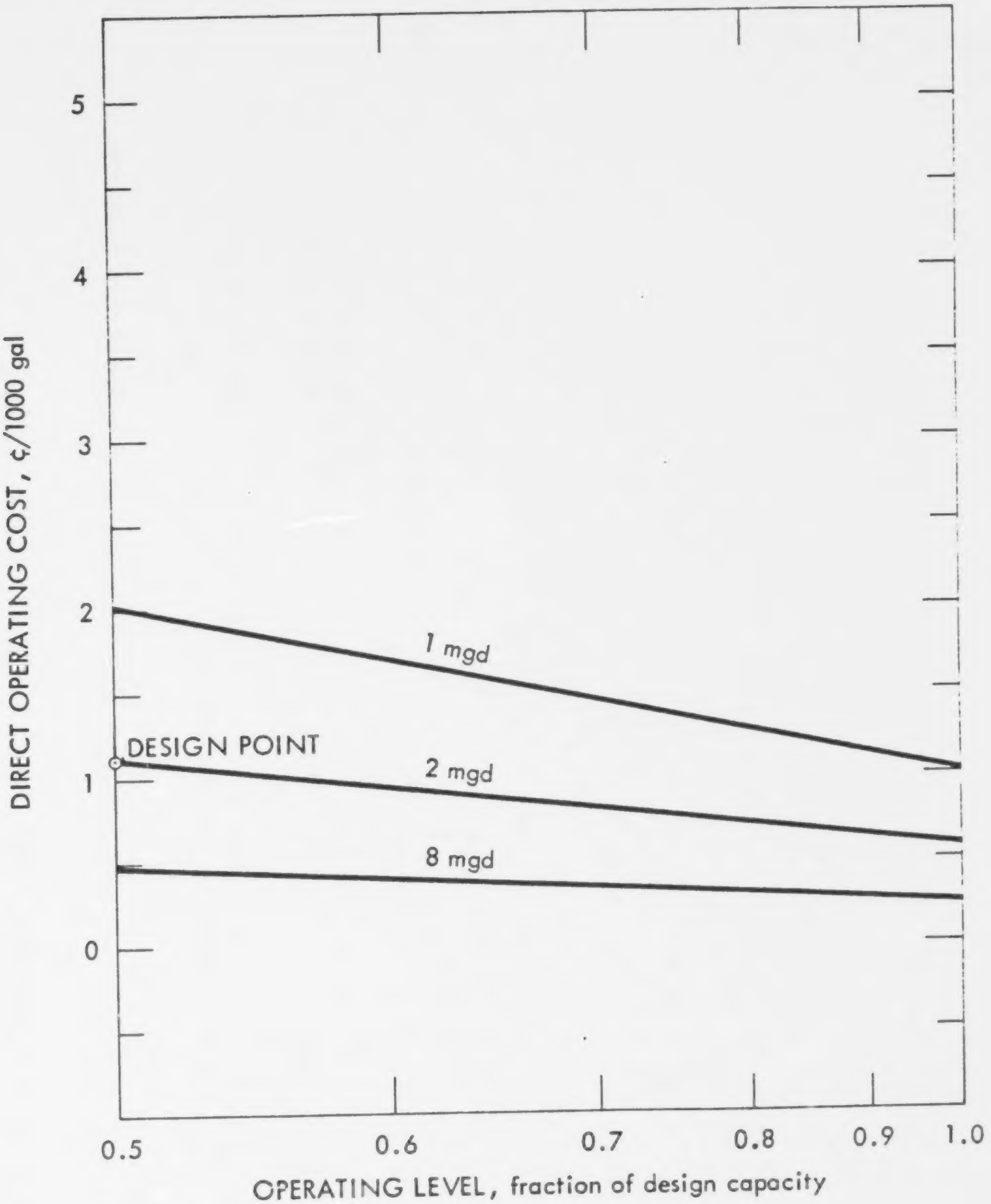


Figure 8-14. CPI separator for 150μ drops: effect of operating level and plant capacity on operating cost.*

*From Item 2 in reference list

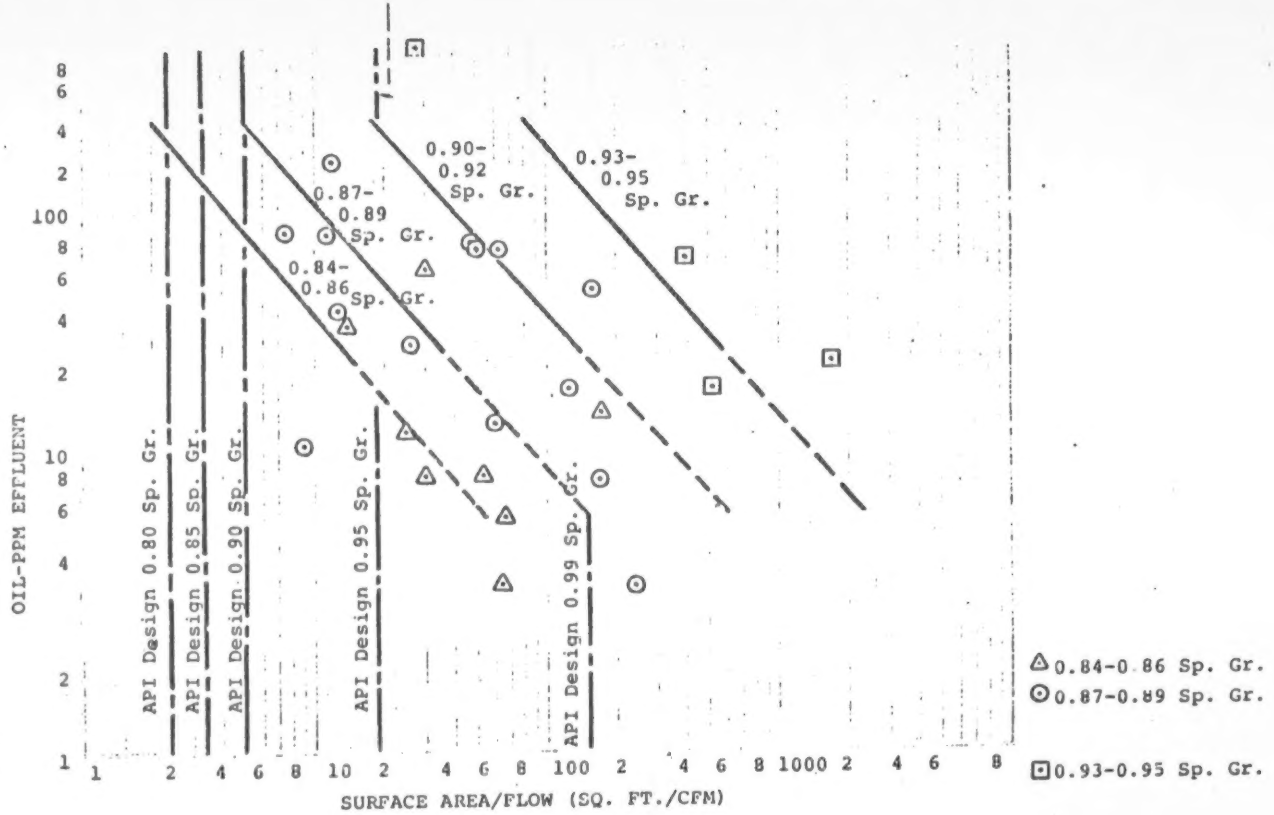


Figure 8-15. Correlation of effluent oil content with operating factors.

*From Item 3 in reference list

ratio, with oil specific gravity as a parameter. Calculated API designs are indicated for several area-to-flow ratios by vertical dashed lines. Although Figure 8-15 was developed from API separator data, a CPI claim of 25 mg/l oil in the effluent for a 0.094 MGD unit checked the predicted figure of 30 mg/l quite well.

Costs--

Thompson concluded that Figure 8-15 applies to separators of any shape and he developed the cost curve relationship for the capital costs of API separators versus required effluent oil concentration that is shown in Figure 8-16. For a reduction from 150 to 10 to 15 mg/l in the effluent, the additional API separator area required was 432,000 square feet, valued at $1.55 million (Oct. 73). In contrast, an air flotation unit was stated to cost $270,000 for the same job.

References--

1. "Manual on Disposal of Refinery Wastes, Volume on Liquid Wastes," American Petroleum Institute, 1969, chp 5. 803*

2. "Treatment of Petrochemical Wastewaters," E.D. Oliver, Stanford Research Institute Process Economics Program Report, No. 80, September 1972. 804*

3. "Data Improves Separator Design," Thompson, S.J., Hydrocarbon Processing, Oct. 1973. 882*

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*Pullman Kellogg Reference File number

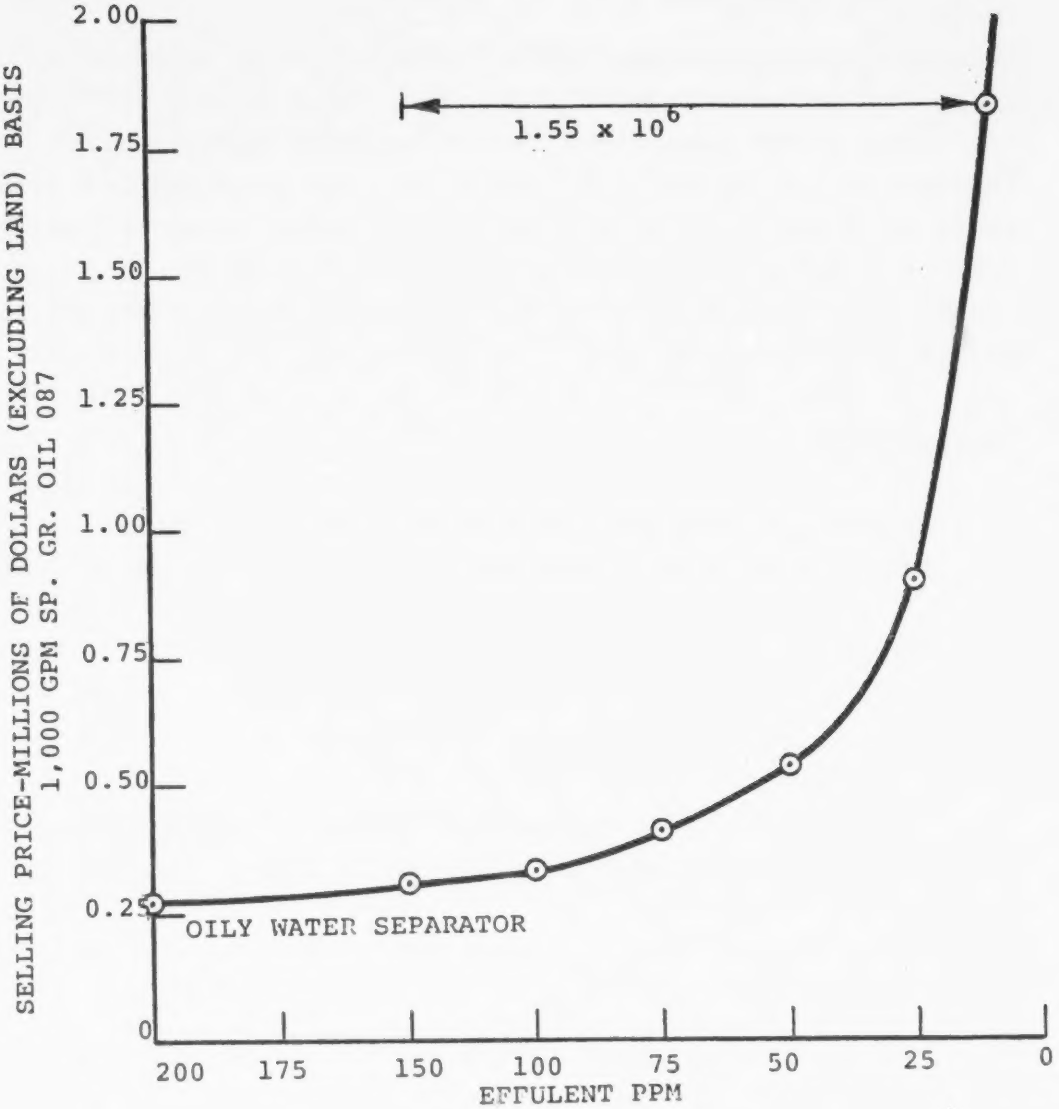

Figure 8-16. Study comparison of covered oily water separator With air flotation.
(From Item 3 in reference list)

Phenol Extraction

The Phenosolvan process typifies phenol extraction methods. It is a liquid-liquid extraction process which removes phenols from aqueous liquors. The process was developed by Lurgi and has been commercial since 1940.

A diagram of the Phenosolvan process is shown in Figure 8-17 (2*). The aqueous liquor feed is first filtered through a gravel bed to remove solids, free tar, and oil. The filtered liquor then passes through a multistage extractor countercurrent to the isopropyl ether solvent. Dephenolized liquor leaves the extractors and may be gas stripped to remove residual solvent. The phenol-rich solvent is first distilled, then stripped to remove the crude phenols and finally recycled to the extractors.

Capability/Efficiency/Limitations--

The Phenosolvan process removes monohydric phenols as well as polyhydric phenols and other organics. Extraction recoveries for coal gasification liquors of 99.5 percent for monohydric phenols, 60 percent for polyhydric phenols and 15 percent for other organics have been estimated in the literature (1). For coal gasification liquors, BOD_5 (5-day BOD analyses) reductions of 84 percent have been estimated. The composition of the crude phenols extracted from coal gasification liquors has been assumed in the literature to be, on a water free basis, 95 percent phenols, and 5 percent other organics, with the phenols being 85 percent monohydric and 15 percent polyhydric. It should be noted that the Phenosolvan process is a licensed process, and the licensor (Lurgi) always includes it as an integral part of the Lurgi Gasification process.

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*Item 2 in reference list

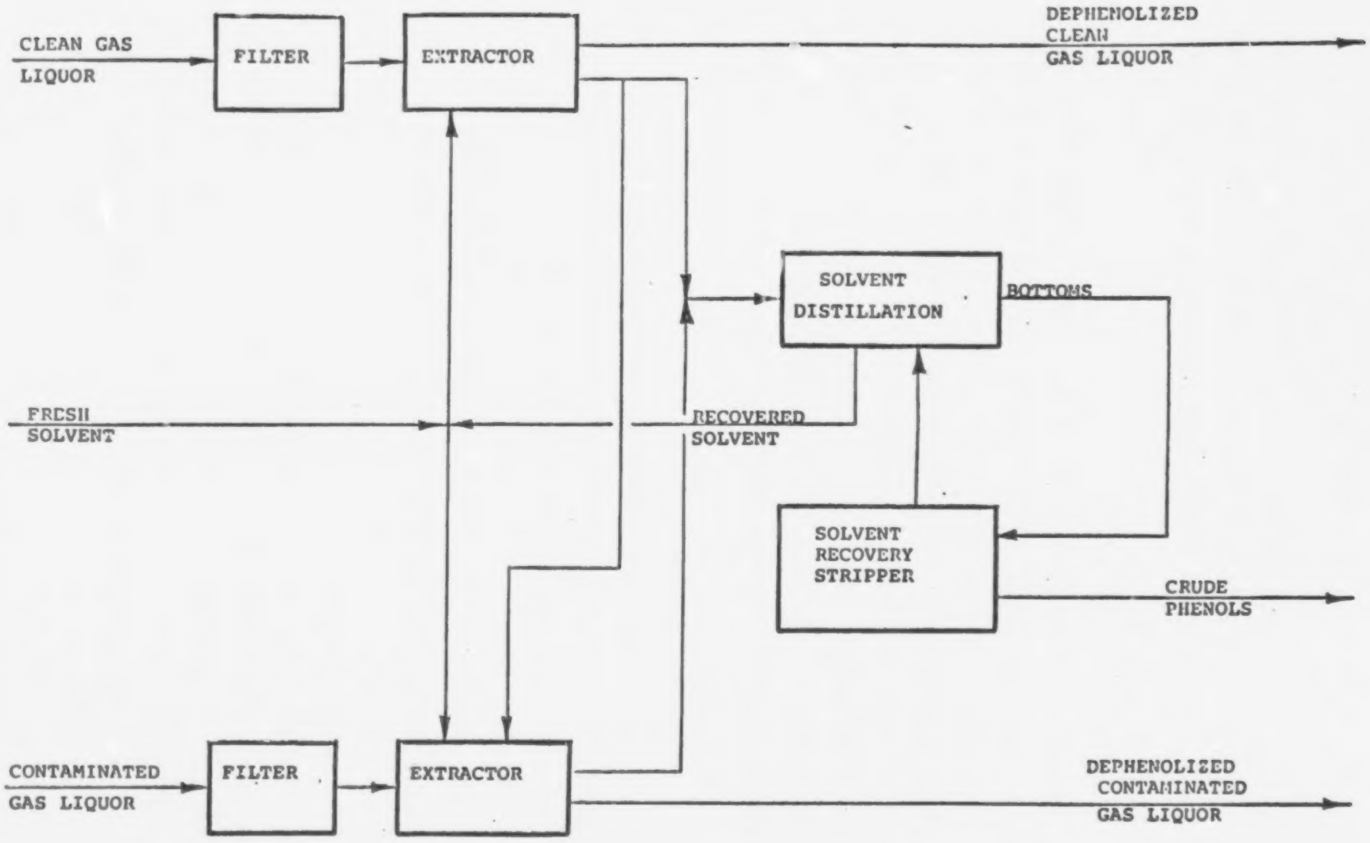
CLEAN GAS
LIQUOR
FILTER
EXTRACTOR
DEPHENOLIZED
CLEAN
GAS LIQUOR
SOLVENT
DISTILLATION
BOTTOMS
FRESH
SOLVENT
RECOVERED
SOLVENT
SOLVENT
RECOVERY
STRIPPER
CRUDE
PHENOLS
CONTAMINATED
GAS LIQUOR
FILTER
EXTRACTOR
DEPHENOLIZED
CONTAMINATED
GAS LIQUOR

Figure 8-17. Phenosolvan process.

(From Item 2 in reference list)

Case Histories--

The Phenosolvan process was originally developed to dephenolize coke oven liquors. Since 1940, 32 commercial units have been built ranging in size from 2 to 1000 GPM of aqueous liquor throughput.

Costs--

American Lurgi has communicated capital and operating costs to us by telephone in response to our letter request. These data appear in a later section entitled "Information Received from Licensors and Vendors".

Possible Improvements--

The use of alternate solvents is the most obvious area for possible improvement.

References--

1. "Coal Gasification and the Phenosolvan Process," Milton R. Beychok, 168th ACS Conference, Sept. 1974, Vol. 19, No. 5, Division of Fuels Chemistry. 830*

2. "Evaluation of Background Data Relating to New Source Performance Standards for Lurgi Gasification," Cameron Engineers Inc., EPA-600/7-77/057, June 1977. 552*

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*Pullman Kellogg Reference number

Stripping and Ammmonia Recovery

Stripping process design depends on the sour water source. The sour waters generated by gasification are high in CO_2, whereas those generated by the liquefaction dissolver contain little or no CO_2. Some gasifier and liquefaction waters contain phenols and phenols interfere with ammonia separation. This description assumes phenol extraction has removed simple phenols.

After the removal of phenols, sour waters containing H_2S, HCN, CO_2, and NH_3 can be steam stripped to remove these gases. High pH stripping favors NH_3 removal and low pH favors H_2S and HCN removal. If steam stripping is carried out in two stages with clean water reflux to the first stage, it is possible to take H_2S and CO_2 overhead from the first stage and recover NH_3 overhead from the second stage. Two-stage steam stripping will provide good H_2S-NH_3 separation only if adequate CO_2 is present.

Because the proprietary United States Steel (USS) Phosam process is said by some sources to be more economical than two-stage stripping, and does not require the presence of CO_2, it appears to be the process of choice in conceptual designs for recovery of H_2S and NH_3.

The proprietary Chevron WWT two-stage stripping process may be desirable for installations producing CO_2-laden sour waters.

Steam Stripping--

Gasification processes producing p/o/t like Synthane (1) and liquefaction processes like SRC (1) and H-Coal (2) may have H_2S concentrations in sour waters in the ranges of 1000 mg/l and 10,000 to 15,000 mg/l, respectively. Both p/o/t gasification and liquefaction processes may have NH_3 concentrations in sour waters in the range of 8,000 to 14,000 mg/l, as shown in TABLE 8-18.

TABLE 8-18. CHARACTERISTICS OF BY-PRODUCT COKE PLANT WASTES. NET PLANT RAW WASTE LOAD*

	EPA PLANTS				Plant	Plant
Characteristics	I	II	III	IV**	A	B
Flow, liters per metric ton coke	580	530	154	19,200	390.9	609.2
Ammonia, mg/l	1,900	1,380	7,330	39	143	110
BOD, mg/l	1,500	1,280	1,120	12	4,140	2,050
Cyanide, mg/l	1,020	110	91	7.7	4.6	3
Oil & Grease, mg/l	--	240	101	2.1	25.3	
Phenol, mg/l	450	350	910	6.1	1,160	430
Sulfide, mg/l	--	1,380	187	4.2		
Suspended Solids mg/l	--	36	421	23		341

* Item 3 in reference list.

**Concentrations are low due to the addition of the final once-through cooler stream which contained significant cyanide.

Coke oven gas liquors and sour waters are usually stripped of H_2S and NH_3 before wastewater treatment. However, as shown in TABLE 8-18, two unstripped waters from plants without once-through cooling dilution were 187 and 1380 mg/l H_2S and three unstripped waters varied from 1,380 to 7,330 mg/l NH_3. Refinery wastewaters are reported to range from 50 to 10,000 mg/l H_2S and 50-7000 mg/l NH_3 (4); however, these figures reflect refinery composite wastewaters, not just sour waters alone. TABLE 8-19 summarizes conventional refinery sour water/composite wastewater stripping experience.

Except for Lurgi gasifiers, few coal conversion wastewater stripping operations are reported in the literature. Lurgi operators can expect 98 to over 99 percent NH_3 removal with a feed containing 11,000 to 16,000 mg/l NH_3 and a 200 mg/l residual (6). Modern ammonia stills at coke liquor operations are designed for a residual of 50 mg/l NH_3 (7) or 96.5 to over 99 percent removal. Conventional oil refinery single-stage steam strippers remove 90 to 97 percent of the NH_3 and 98 to 99 percent of the H_2S from refinery sour waters (4).

The proprietary two-stage Chevron WWT process is designed to recover NH_3 and H_2S separately and has residuals of <50 mg/l NH_3 and <5 mg/l H_2S (8).

Stripping costs and a sketch of this process were provided by Chevron Research and are included in a later section entitled "Budget Cost Estimates Received From Process Licensors". Ammonia recovery is included in this presentation.

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*Item 1 in reference list

TABLE 8-19. REFINERY SOUR WATER STRIPPER OPERATION*

Example:	1A**	1B	2A	2B	3	4	5	6	7	8A	8B
CONDITIONS											
Raw feed											
Flow, gpm	150	200	80	60	204	35	89	41	250	125	258
Temperature, deg F						113		100	240	190	190
Hydrogen sulfide, ppm	10,000	8,000	1,500	1,876	5,000	4,600	3,000	7,350	2,000	2,180	3,120
Ammonia, ppm	5,000	5,000	1,000	1,480	5,000	3,800	1,500	9,800	5,000	1,430	1,560
Reflux											
Flow, gpm	12	33	none	none	10	3	15	7.7	none	none	none
Temperature, deg F	200				200	226		140			
Tower feed †											
Flow, gpm	182	233	60	80	214	38	100	48.3	250	325	258
Temperature, deg F	200	200	195	195	240	160	150	207	240	190	190
Tower bottoms											
Flow, gpm	195	250	62	62	216	41	105	52.7	255	344	274
Temperature, deg F	268	268	230	230	250	240	220	240		240	240
Pressure, psig	30	22	6	6-7		15		10	11		
Hydrogen sulfide, ppm	50	10	2	0	50	30	5	trace	5	32	36
Ammonia, ppm	280	200	300	194	600	227		700	200	580	490
Stripping steam, lb per hour	15,300	16,700	1,400	2,400	4,870	1,400	4,700	4,500	10,000	7,000	9,200
Stripping steam, lb per gallon ‡ §	1.41	1.2	0.39	0.67	0.38	0.62	0.78	1.54	0.67	0.36	0.60
Hydrogen sulfide removal, percent	99.4	99.9	99.9	100.0	98.9	99.0	99.8	100.0	99.75	98.3	99.0
Ammonia removal, percent	94.0	95.0	69.0	86.5	87.3	93.0		89.6	96.0	57.3	67.0
Trays (or packing)	13	13	(12 ft)	(12 ft)	6	6	12	12	10	4	4
References:	1,5	1,5	1,5	1,5	1,5,6	1,5,7	1,8	1,9	10	11	11

Example:	9A	9B	10A	10B	10C	11	12	13A	13B	14	15	16
CONDITIONS												
Raw feed												
Flow, gpm	80	80	75	87	150	95	40	33	26	75	148	135
Temperature, deg F	110	110	100	100	100	100	100	85	85			
Hydrogen sulfide, ppm	330	340	6,550	12,010	5,600	3,830	5,500	600	2,910	7,200	5,000	1,810
Ammonia, ppm	1,310	1,310	3,700	3,620	2,640		5,500	2,640	4,620	5,000		1,769
Reflux												
Flow, gpm	none	none	21	9	3	none	none	none	none	none	none	17
Temperature, deg F			180	180	175							215
Tower feed †												
Flow, gpm	80	80	96	96	153	95	40	33	26	85	148	152
Temperature, deg F	110	110	210	210	210	100	250	85	85	220	200	190
Tower bottoms												
Flow, gpm	89	89	98	98	156	106	40	38	90	86	151	158
Temperature, deg F	215	215				214	260			230		232
Pressure, psig	7	7	7.5	7.5	8.5					4.7		
Hydrogen sulfide, ppm	0	0	16	19	18	380	50	115	0.3	238	206	1
Ammonia, ppm	124	64	47	78	128		750	14	120			90
Stripping steam, lb per hour	2,200	5,500	14,500	14,000	11,900	550	2,100	2,100	970	1,550	2,500	9,300
Stripping steam, lb per gallon ‡ §	0.46	1.13	2.52	2.43	1.30	0.10	0.87	1.06	0.62	0.3	0.28	1.02
Hydrogen sulfide removal, percent	100.0	100.0	99.7	99.8	99.7	89.0	99.0	100.0	100.0	96.2	95.8	99.9
Ammonia removal, percent	89.0	95.0	98.3	97.6	95.0		86.4	98.5	97.0			98.0
Trays (or packing)	(15 ft)	(15 ft)	(15 ft)	(15 ft)	(15 ft)	8	(20 ft)	(15 ft)	(15 ft)	6	8	8
References:	4	4	4	4	4	4	4	4	4	1,5	4	11

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* Item 4 in reference list

** A, B, and C examples are different operating conditions for the same stripper

† Tower feed is raw feed plus reflux

‡ Pounds of steam per gallon of tower feed

§ Does not include the heating steam required to heat the tower feed to the tower bottoms temperature

David G. Rodriguez' article on Atlantic Richfield stripping experience (9) is interesting for its general design description and, particularly, its insights into system design for reliable operation. Rodriguez states that the gathering system should (a) provide good oil/water interface controls at each sour water source, (b) have adequate feed equalization to prevent hydrocarbons from entering the stripper, (c) provide good skimming in the feed equalization and (d) smooth or absorb feed pressure variations in feed equalization.

The stripper equipment should be designed for the flow and quality of composite feed expected and the desired concentrations of H_2S and NH_3 in the stripped waters. Desired composition of the stripped waters will depend on the downstream process sequence. Consideration must be given to operations with varied NH_3 residual in the stripper bottoms.

Stripper downtime other than that due to hydrocarbon carryover is caused by corrosion and salt deposits. Careful materials selection reduces corrosion. Salt deposits are minimized by maintaining temperatures above 150°F.

The stripper overhead gas may be processed for separation and recovery of ammonia, flared,or incinerated.

Stripping Applications--

Stripping H_2S results in 98 to 99 percent recovery of the sulfur values in some waters without exceeding biological oxidation (biox) sulfur nutrient requirements, in-plant recycle requirements, or effluent discharge limitations. Stripped waters may contain 0 to 10 mg/l residual H_2S. Biox requirements are about (1/3,000) x BOD or 5 to 10 mg/l H_2S for p/o/t gasification or liquefaction waters. The remaining H_2S is oxidized to sulfates before reuse or discharge. Treating processes for cooling tower sidestreams or boiler feed waters can handle the residual sulfates. The most stringent U.S. sulfate discharge regulation is 250 mg/l and is readily met.

By stripping, ammonia can be removed from wastewaters down to biox nutrient requirements but not to recycle or most stringent discharge requirements. Stripping can produce wastewaters containing 50 mg/l NH_3 or less. Biox nitrogen requirements are about (1/33) x BOD or (1/14) x phenol, thus about 100 to 1,000 mg/l NH_3 is needed for undephenolized wastes. Ammonia stripping prior to biox treatment must be controlled to leave enough nitrogen for the phenol/BOD load or NH_3 from NH_3recovery returned to supply adequate nutrients. Biox removes the nitrogen (ammonia or nitrate) needed for biomass growth and good biomass separation, and subsequent treatment should produce an effluent with NH_3 sufficiently low to meet either recycle or discharge limits.

Gasification processes producing no p/o/t with low biox potential require special handling after conventional ammonia stripping. This special handling can be simple second-stage ammonia stripping and/or breakpoint chlorination or other treatment. Recent articles from American Petroleum Institute (12, 13) have an excellent discussion of "ammonia fixation" in refinery strippers, which explains the variability of performance shown in TABLE 8-19. Ammonia can be freed for stripping by addition of lime or caustic to pH 9.5 to 11.

Case History of Atlantic Richfield Stripping Experience--
The Atlantic Richfield stripper embodies the principles described in Rodriguez' article (9). Its operation reduced combined refinery waste COD from 277 to 114 mg/l with (calculated) composite H_2S and NH_3 sour water concentrations of 5,360 and 1,850 mq/l, respectively. Thus, although sour waters may be only 5 to 15 percent of the combined refinery water flow, stripping sour water reduced refinery COD almost 60 percent.

Case History of Chevron WWT Process--

Chevron's WWT process has been operating in Richmond, Cal. since 1969. It recovers NH_3 and H_2S in two stages similar to conventional two-stage steam stripping with most of the H_2S and associated CO_2 removed in the first stage (8). The residuals from the second stage are <50 mg/l NH_3 and <5 mg/l H_2S and are suitable for refinery reuse. An ammonia scrubbing system is required on the second stage overheads to remove H_2S and other impurities.

Phosam-W--

Phosam-W is US Steel's designation for an adaptation of their Phosam process to coal gasification and other non-coke oven processes.

The Phosam-W process requires three major vessels, the "superstill," the ammonium phosphate stripper and the ammonia fractionator. Figure 8-18 is a simplified block flow diagram of the Phosam-W process (1). In operation, sour water feed to the superstill is heated to the bubble point and fed to the top of the stripper (lower) section of the superstill at 2.8 N/cm^2 (4 psig).

The superstill is heated indirectly with 41.4 N/cm^2 (60 psig) steam at 150°C and, at 2.8 N/cm^2 (4 psig), the stripped water contains less than 200 mg/l free NH_3. The upper section of the superstill is an ammonium phosphate absorber with a spray section topped by a tray column. Acid vapors from the stripper section of the superstill pass upward through a sprays-and-tray absorber and exit at 2.1 N/cm^2 (3 psig). Superstill acid vapors consist largely of H_2S and whatever CO_2 was in the sour water feed. Rich ammonium phosphate from the absorber section is pumped, via heat exchange with the overhead vapors, to the ammonium phosphate stripper.

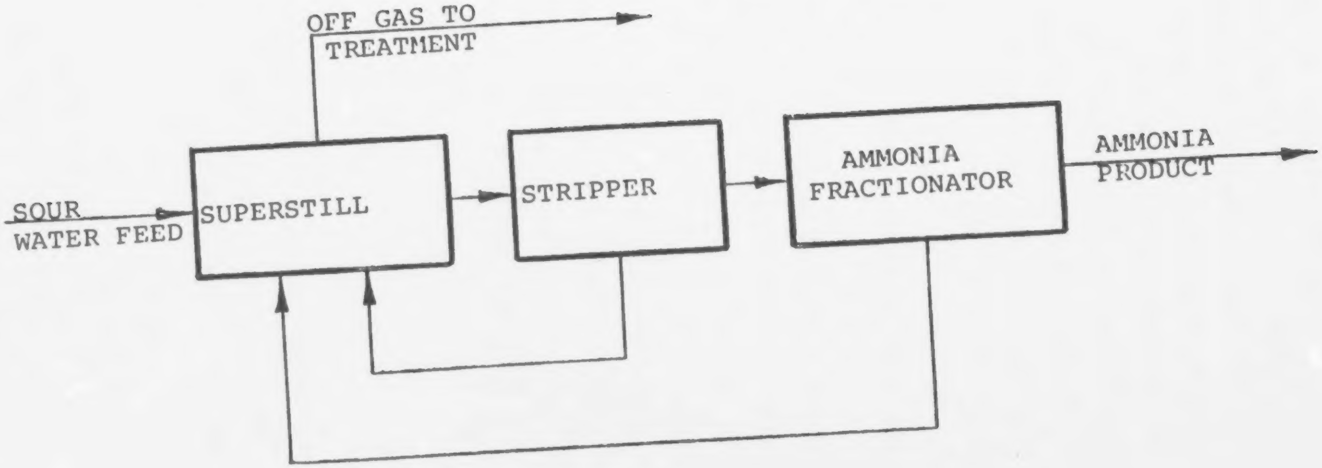
OFF GAS TO
TREATMENT
SOUR
WATER FEED
SUPERSTILL
STRIPPER
AMMONIA
FRACTIONATOR
AMMONIA
PRODUCT

Figure 8-18. Phosam-W process for ammonia separation.
(From Item 1 in reference list)

The ammonium phosphate stripper is operated at elevated pressure with heat provided by 414 N/cm^2 (600 psig) steam at 249°C to strip NH_3 from the rich ammonium phosphate. The hot lean ammonium phosphate is cooled before entering the superstill absorber as the scrubbant. The stripper vapors, containing 10 to 20 percent NH_3 in water vapor, are condensed with heat recovery and enter the ammonia fractionator feed tank. The ammonia fractionator column strips anhydrous NH_3 from the water and is operated at elevated pressure. Product NH_3 leaves the top of the fractionator via heat exchange with cooling water and refluxes to the column. The hot water from the bottom of the fractionator is flashed in the bottom of the superstill to provide the rest of the required heat and for recovery of some NH_3. The ammonia fractionator bottoms may contain up to 500 mg/l NH_3 (10).

The NH_3 content of the stripped wastewaters depends on the design of the stripper section of the superstill but has been 100 to 200 mg/l (10) in designs up to 1975.

Utility requirements for Phosam-W are given in TABLE 8-20. The unusual features of the process are that it will absorb ammonia out of a mixture of acid gases such as H_2S, CO_2 and HCN without co-absorption of acid gas and that it can operate on liquid, gas, or vapor feeds.

References--

1. EPA-600/7-77-065 "Water Conservation and Pollution Control in Coal Conversion Processes". Water Purification Associates. June 1977. P. 257, 260, 265, 267. 480*

2. Reap, E.J., Davis, G.M., Duffy, M.J., and Koon, J.H., "Wastewater Characteristics and Treatment Technology for the Liquefaction of Coal Using H-Coal Process," Proceedings of the 32nd Purdue Industrial Waste Conference, May 1977. 654*

TABLE 8-20. UTILITY REQUIREMENTS FOR A TYPICAL PHOSAM-W PLANT INCLUDING WASTEWATER STRIPPING*

	Per Pound of NH_3
Steam at 550 psig, lb	12
Steam at 25 psig, lb	8
Cooling Water, gal	40
Electrical Power, KWH	0.03
Chemicals	
H_3PO_4 Makeup (as 100% H_3PO_4), lb	0.002
NaOH (as 100% NaOH), lb	0.003

*From Item 1 in reference list

3. Adams, C.E., Jr., Stein, R.M., and Eckenfelder, W.W., Jr., "Treatment of Two Coke Plant Wastewaters to Meet Guideline Criteria", Proceedings of 29th Purdue Industrial Waste Conference, May 1974. P. 864-880.

4. "Manual on Disposal of Refinery Wastes, Volume on Liquid Wastes", American Petroleum Institute, 1969. Chp. 10. 803*

5. Sylvester, N.C., "Sour Water Treatment, A State-of-the-Art Review". University of Tulsa Technical Report Number 74-1. 641*

6. EPA-600/7-77-057 "Evaluation of Background Data Relating to New Source Performance Standards for Lurgi Gasification". Cameron Engineers, Inc., June 1977. P. 126-129. 552*

7. Parsons, W.A., and Nolde, W., "Applicability of Coke Plant Water Treatment Technology to Coal Gasification". Presented EPA Symposium, Hollywood, Florida, September 1977. 958*

8. Klett, R.J., "Treat Sour Water for Profit", Hydrocarbon Processing. October 1972.

9. Rodriguez, D.G., "Sour Water Stripper: Its Design and Application". AIChE Symposium Series: Water-1973, No. 136, Vol. 70. 774*

10. FE-1772-11 (ERDA) "Handbook of Gasifiers and Gas Treatment Systems", Dravo Corporation, February 1976. P. 142-144. 266*

11. Armstrong, T.A., "There's Profit in Processing Foul Water". Oil and Gas Journal, June 1968.

12. Gantz, R.G. "API-Sour Water Stripper Studies", 40th Midyear meeting of Division of Refining, API, Chicago 1975. Preprint No. 0375.

13. Bomberger, D.C., and Smith, J.H., (Stanford Research Institute) "Evaluation of Ammonia Fixation in Actual Refinery Sour Waters", API Report 954, January 1978. *843

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*Pullman Kellogg Reference File number

Coagulation and Flocculation

Coagulation and flocculation (C&F) are, respectively, chemical and physical means of clarifying wastewater and are usually used in conjunction with either flotation or filtration. Flotation and wastewater filtration are described separately.

Coagulation and flocculation occur by a combination of chemical and physical factors (1, 2*) and may or may not be accompanied by chemical precipitation. Coagulation is used for removing suspended solids or colloidal particles from water and precipitation is used for removing some dissolved solids, mainly hardness and trace metals. After coagulation and/or precipitation, flocculation follows before subsequent solids removal.

Coagulation requires the destabilization of colloids, usually by charge modification coupled with bridging or enmeshment processes (1). The two principal means of determining optimum doses for C&F are the jar test and zeta potential measurement. The jar test is a lab method of C&F followed by sedimentation. For the usual jar test, coagulant is added to wastewater at several dose levels and/or floc times using six stirred jars. Mixing and flocculation are followed by a settling period after which samples of supernatant liquid are examined for turbidity. Figure 8-19 shows the results of tests at several doses and settling times with ferric sulfate. Doses and times are chosen to provide desired effluent quality at least cost. Zeta potential requires a special apparatus that measures the velocity of floc migration across a fixed electrical potential field. The zeta potential for best C&F results is used for process control. Figures 8-20(a) and 8-20(b) show the zeta potential and residual turbidity, respectively, as functions of alum (aluminum sulfate)

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*Items 1 and 2 in reference list

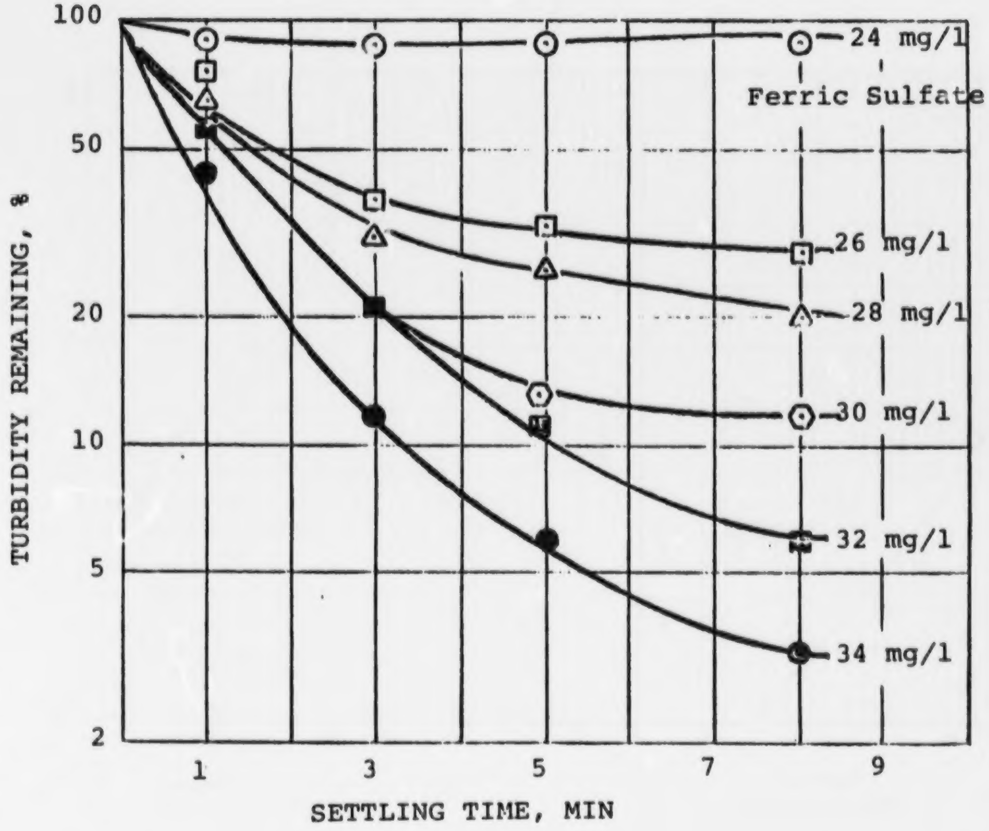

Figure 8-19. Jar test results.*
*From Item 1 on reference list

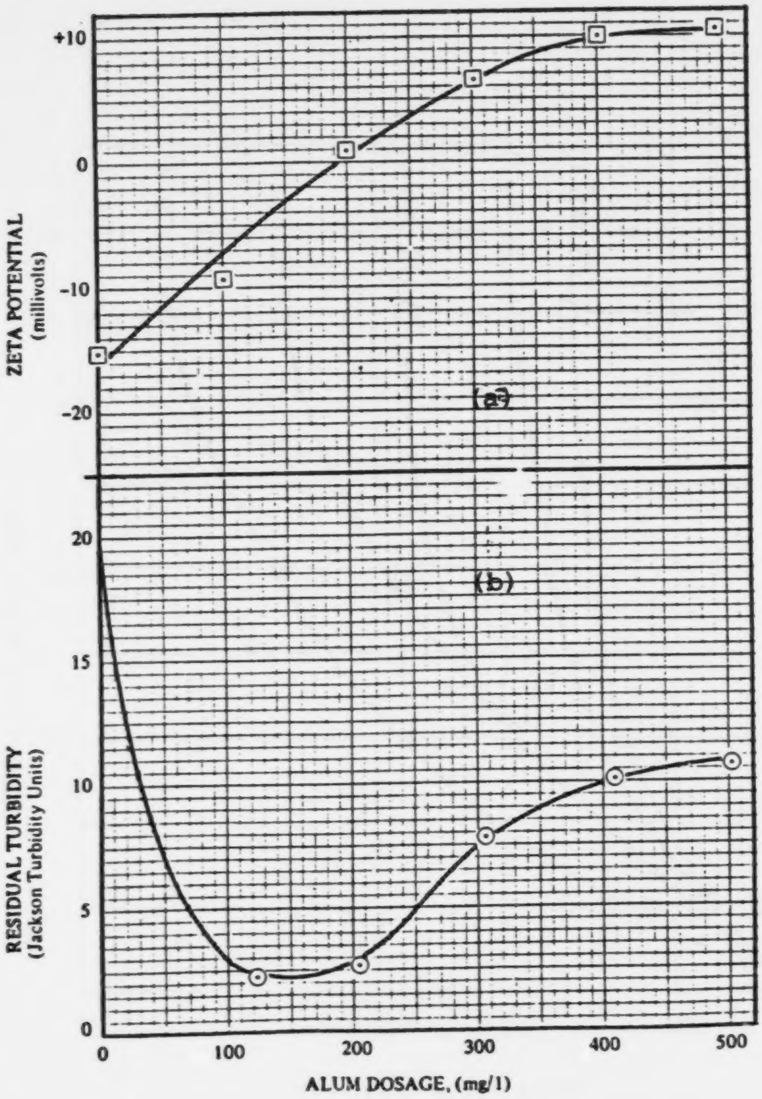

Figure 8-20. Coagulation of Raw Sewage with Alum.*

*From Item 1 on reference list

dose. The zeta potential, Figure 8-20(a), corresponding to the best alum dose in Figure 8-20(b) is -2 to -4 millivolts.

Charge modification can result in restabilization if coagulant overdose occurs. Figure 8-20(b) shows the effects of alum overdose as turbidity passes through a minimum at 150 mg/l then begins to rise at higher alum rates.

As the solubilities of the chemicals added for treatment are exceeded, precipitation occurs and the precipitates form flocs. The usual C&F chemicals are alum, ferric or ferrous sulfate,and lime. The precipitation reactions depend on the chemical treatment and the water characteristics. TABLE 8-21(a) summarizes the important reactions of alum in water treatment. Generally CO_2 is released from the waters and alkaline carbonates and aluminum hydroxide, $Al(OH)_3$, precipitate. Various alkalis may be used to maintain pH. TABLE 8-21(b) shows the similar reactions of ferric sulfate, $Fe_2(SO_4)_3$. TABLE 8-21(c) shows that ferrous sulfate reacts with only 1/3 the amount of lime of either alum or ferric sulfate. TABLE 8-21(d) summarizes the reactions of lime with calcium and magnesium hardness in the wastewater (as bicarbonates) and with soda ash added to increase pH (alkalinity) and calcium carbonate precipitation. Note that only insoluble carbonates form and gases are not evolved.

In addition to the usual raw water coagulants (alum, lime, and the iron salts) other compounds may be used as supplements or enhancers. Flocculating agents, i.e., polymers or polyelectrolytes, may be added to alum or iron flocs to increase floc size and strength. Soda ash may be added to increase alkalinity for lime processing while a strong acid or base may be used for pH adjustment.

TABLE 8-21(a). REACTIONS OF ALUMINUM SULFATE*

$$Al_2(SO_4)_3 + 3\,Ca(HCO_3)_2 \rightarrow 2\,Al(OH)_3\downarrow + 3\,CaSO_4 + 6\,CO_2\uparrow$$

$$Al_2(SO_4)_3 + 3\,Na_2CO_3 + 3\,H_2O \rightarrow 2\,AL(OH)_3\downarrow + 3\,CO_2\uparrow$$

$$Al_2(SO_4)_3 + 3\,Ca(OH)_2 \rightarrow 2\,Al(OH)_3\downarrow + 3\,CaSO_4$$

TABLE 8-21(b). REACTIONS OF FERRIC SULFATE*

$$Fe_2(SO_4)_3 + 3\,Ca(HCO_3)_2 \rightarrow 2\,Fe(OH)_3\downarrow + 3\,CaSO_4 + 6\,CO_2\uparrow$$

$$Fe_2(SO_4)_3 + 3\,Na_2CO_3 + 3\,H_2O \rightarrow 2\,Fe(OH)_3\downarrow + 3\,Na_2SO_4 + 3\,CO_2\uparrow$$

$$Fe_2(SO_4)_3 + 3\,Ca(OH)_2 \rightarrow 2\,Fe(OH)_3\downarrow + 3\,CaSO_4$$

TABLE 8-21(c). REACTIONS OF FERROUS SULFATE*

$$FeSO_4 + Ca(HCO_3)_2 \rightarrow Fe(OH)_2\downarrow + Ca\,SO_4 + 2CO_2\uparrow$$

$$FeSO_4 + Ca(OH)_2 \rightarrow Fe(OH)_2\downarrow + Ca\,SO_4$$

$$4\,Fe(OH)_2 + O_2 + 2H_2 \rightarrow 4\,Fe(OH)_3\downarrow$$

TABLE 8-21(d). REACTIONS OF LIME*

$$Ca(OH)_2 + Ca(HCO_3)_2 \rightarrow 2\,CaCO_3\downarrow + 2H_2O$$

$$2\,Ca(OH)_2 + Mg(HCO_3)_2 \rightarrow 2\,CaCO_3\downarrow + Mg(OH)_2\downarrow + 2H_2O$$

$$Ca(OH)_2 + Na_2CO_3 \rightarrow CaCO_3\downarrow + 2\,NaOH$$

*From Item 1 in reference list

Alum and iron salts release the appropriate cation, Al^{3+} or Fe^{3+}, and these form insoluble hydroxide flocs whose charge depends on the pH. At or near the floc isoelectric point (near neutral for alum and iron) flocs aggregate and settle. Alum's solubility as hydroxide is also pH dependent in the range of interest and it passes through a minimum near pH 7 (2). Lime acts to remove suspended solids by precipitating calcium carbonate, $CaCO_3$, and to do so carbonate or bicarbonate alkalinity must be present. Fortunately the high pH required for effective lime coagulation (9.5-11.5) also helps to remove many heavy metals. These metals will precipitate with the lime sludge. Figures 8-21 and 8-22 show, respectively, the optimum pH values for various metal removals singly and in the presence of ammonia. See "Chemical Precipitation" for further discussion of lime treatment to remove calcium and magnesium hardness and various heavy metals.

Chemicals Required--

Quantities of chemicals required for C&F treatment depend on the characteristics of the wastewater, including pH, temperature, hardness,and concentration of suspended solids. With reference to equations in TABLES 8-21(a) and 8-21(b), there are the following relationships for chemical equivalency:

ALKALINITY EQUIVALENCY FOR CHEMICAL TREATMENTS*

	Chemical Treatment Addition of 1 mg/l		
Reactant Equivalent	Alum	Ferric Sulfate	Ferric Chloride
Alkalinity, as $CaCO_3$	0.50	0.57	0.92
95% Hydrated lime as $Ca(OH)_2$	0.39	0.44	0.72
Soda ash as Na_2CO_3	0.54	0.62	1.00

*Item 1 in reference list.

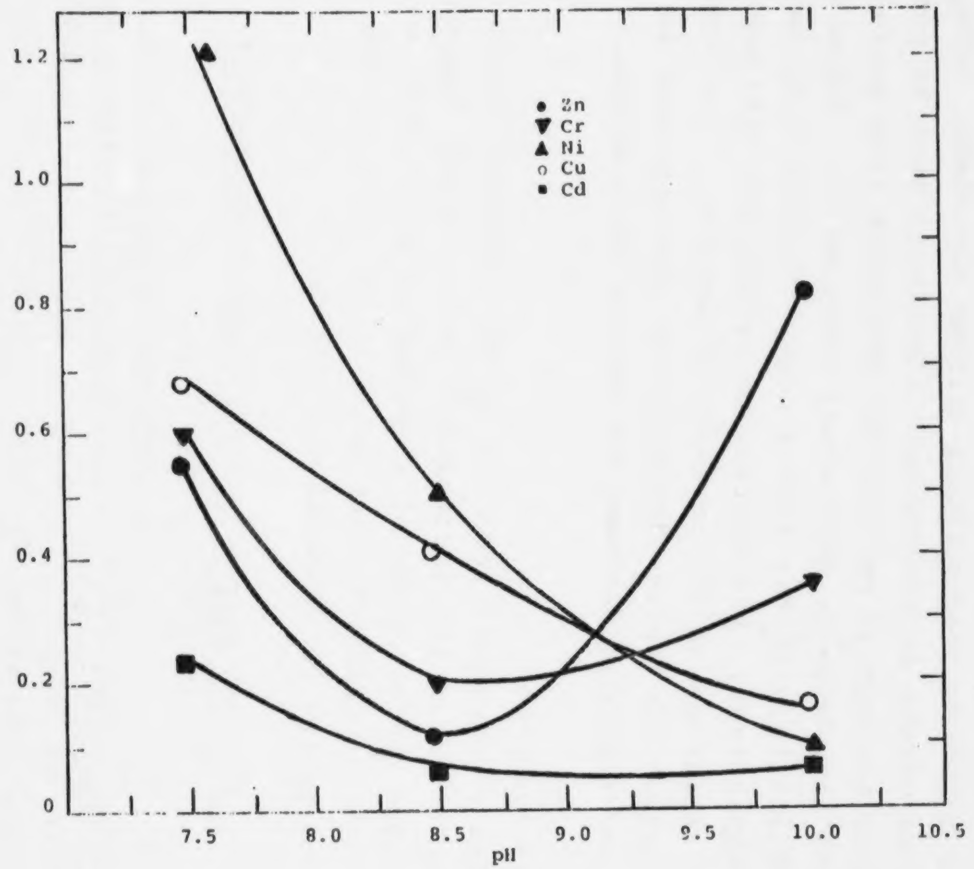

Figure 8-21. Optimum pH values for metal removal.*

*From Item 12 on reference list, p. V-B/125

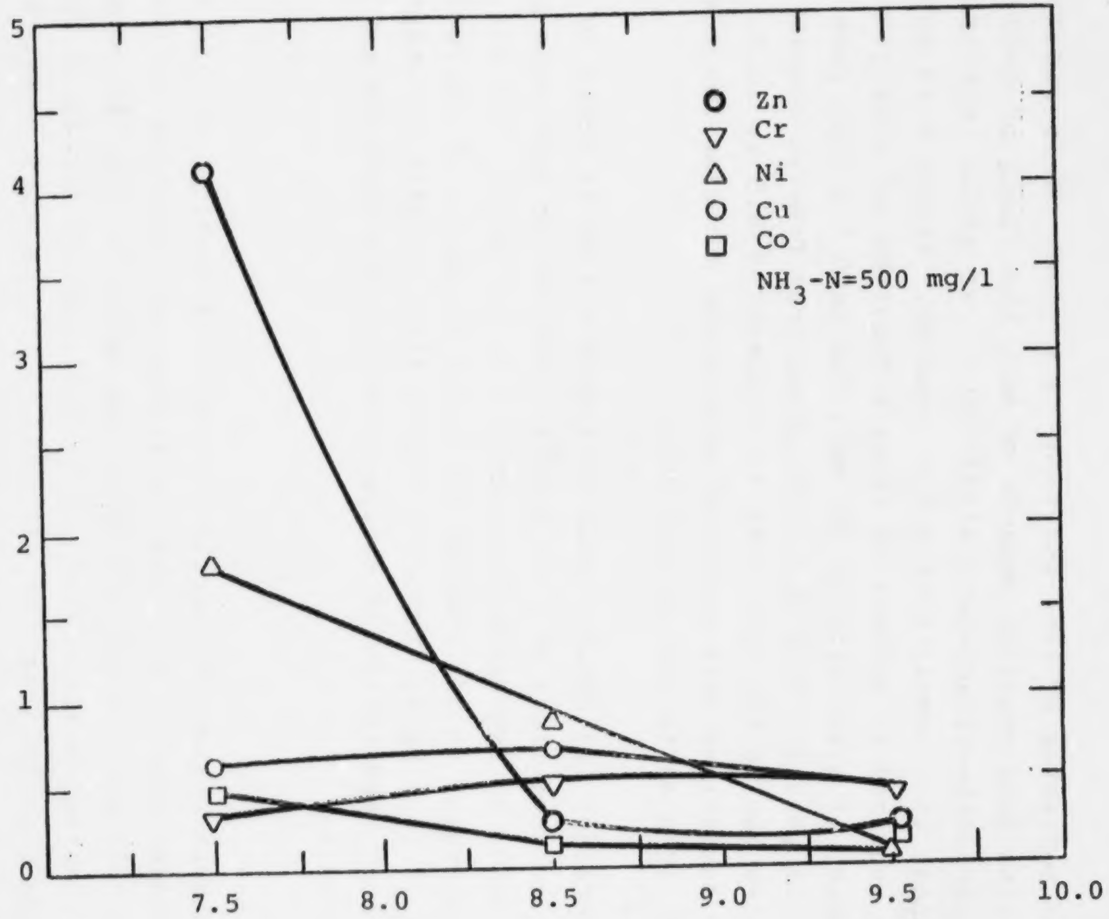

Figure 8-22. Optimum pH values for metals removal in the presence of ammonia*

*From Item 12, p. V-B/125, on reference list

These amounts of alkali will just maintain the alkalinity unchanged when 1 mg/l chemical is added. For example if no alkalinity is added 1 mg/l alum will reduce the alkalinity 0.5 mg/l as $CaCO_3$.

Water softening by lime addition to a given pH is widely used. The lime dose required depends on both the final pH desired and the carbonate-bicarbonate alkalinity. At pH 11 virtually all hardness will precipitate and be removed. Figure 8-23 shows the lime required to achieve pH 11 as a function of alkalinity; for example, if alkalinity is 100 mg/l (as $CaCO_3$) a lime dose of 200 mg/l is needed. Many municipal phosphorus removal operations (1) require 200 to 600 mg/l lime to achieve pH 9.5 to 11.5. TABLE 8-22 summarizes this municipal experience and also includes some alum, iron salts and polymer data.

Eckenfelder (2) reports that activated silica is added to toughen alum or iron flocs at 2 to 5 mg/l. Anionic or non-ionic polymers are added to aggregate flocs at 0.2 to 1.0 mg/l. The EPA report on suspended solids removal (1) gives examples of polymer usage of 0.08 to 1 mg/l. This EPA report also includes in Chapter 5 a thorough description of equipment design for handling and storing chemicals.

Costs--

Treatment costs for coagulation and flocculation include cost of treatment chemicals, costs for storage and handling of treatment chemicals and capital and operating costs for the C&F equipment. The chemical sludges produced by settling, filtering,or flotation require processing for disposal. The related costs will be greater than for non-chemically produced sludges: more sludge is produced, since more suspended solids are removed, and some of the added chemicals also precipitate.

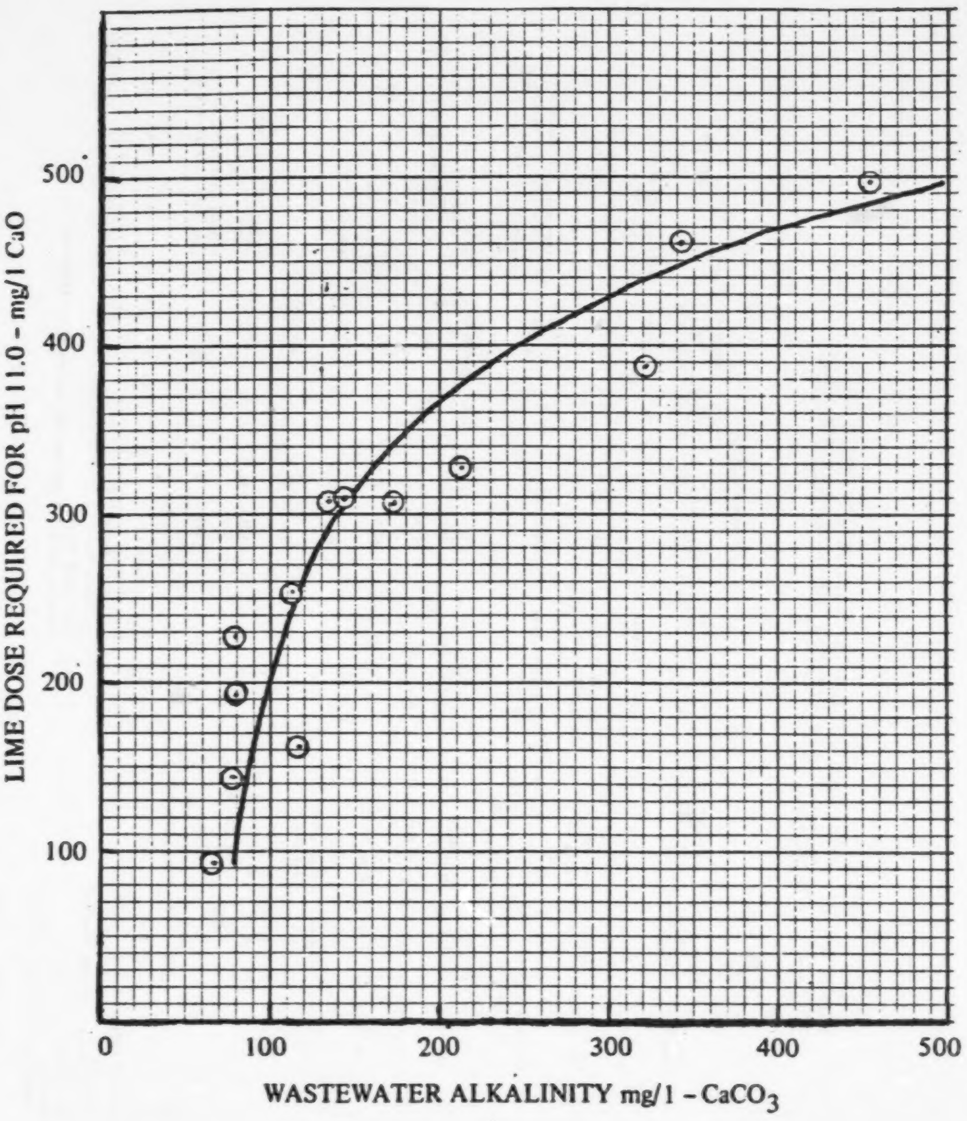

Figure 8-23. Lime requirement for pH> 11 as a function of wastewater alkalinity.*

*From Item 1 in reference list

TABLE 8-22. SUSPENDED SOLIDS REMOVAL PERFORMANCE FOR CHEMICAL COAGULATION APPLICATIONS TO PHOSPHATE REMOVAL*

LOCATION	PROCESS	PLANT SIZE mgd	AVERAGE CHEMICAL FEED mg/l	pH	BASIC EQUIPMENT	SUSPENDED SOLIDS Inf. mg/l	SUSPENDED SOLIDS Settled Eff. mg/l	DATA PERIOD Months	OVERFLOW RATE gpd/sq ft	COMMENTS
Lebanon, Ohio	IPC	0.1	Lime ~250	9.5	1-stage SC	109	30	45	1440	Acid pH adjustment. Reference 19
EPA, Blue Plains Plant, Washington, D.C.	IPC	0.1	Lime 460$^{(a)}$ +$FeCl_3$ 5$^{(b)}$	11.5$^{(a)}$ 10.0$^{(b)}$	2-stage SC(a) FM,FL,S$^{(b)}$	158	14.4	6	500-1800	Two stages; with intermediate recarbonation. Reference 20
Ely, Minn.	Tertiary	1.5	Lime 250-350$^{(a)}$ +Polymer .2$^{(a)}$ +$FeCl_3$ 6$^{(b)}$	11.8$^{(a)}$ 9.5$^{(b)}$	2-stage SC	75	10	5	570$^{(a)}$ 660$^{(b)}$	System designed for nearly complete (eff. ≤ 0.05 mg/l) phosphorus removal. Reference 21
S. Lake Tahoe, California	Tertiary	7.5	Lime 400 275	11.3 10.5	FM,FL, S SR	38 38	10 25	21	400-600	Recarbonation Reference 2[illegible]
Lebanon, Ohio	Tertiary	0.1	Lime 220-270	9.5	SC	43.5	16.5	10	1440	Acid pH adjustment made. Reference 23
Nassau County, New York	Tertiary	0.6	Alum 20	-	SC	22.5	2.5	72	860	Bulking 2-3 times a year. Reference 24
Salt Lake City, Utah	IPC	0.04-0.1	Iron 34-41 Polymer 0-1.5	-	SC	90	21.9	2.5	360-1080	Reference 11
		0.05-0.09	Alum 14 Polymer 0-0.25	-	SC	95	26.9	1	500-870	---
		0.03-0.18	Lime 270-586	9.8-11.0	SC	101	11.6	5	290-1800	H_2SO_4; pH adjustment.
Leetsdale, Pa.		0.6	$FeCl_3$ 100 +Polymer 0.5	6.5-6.7	S	280	38	-	-	Primary Treatment, Constant Feed - Reference 9

Key:
FM - Flash Mix Unit
FL - Flocculator
SC - Solids Contact (Sludge Blanket)
SR - Solids Recirculation
S - Settling
IPC - Independent Physical Chemical

Notes:
Lime as $Ca(OH)_2$
Alum as Al^{+++}
Iron Salts as Fe^{+++}
(a) First Stage
(b) Second Stage
Overflow rates are minimum and maximum where range appears.

*Item 1 in reference list

The costs of chemicals vary with the market but a recent comparison is shown in TABLE 8-23 for the relative costs (1977) of chemicals for pH adjustment. Lime and soda are included in TABLE 8-23 since they are used extensively in coagulation and water softening. Current costs from "Chemical Marketing Reporter" for alum, ferric sulfate, and quicklime are about \$137, \$99, and \$25 to 45, respectively, per dry ton, bulk, FOB factory. Lime is produced locally throughout much of the U.S. and may have lower shipping costs than either alum or ferric sulfate. Thus compared to high calcium quicklime, relative costs for alum and ferric sulfate are respectively about 4X and 2 to 3X.

Total costs are difficult to ascertain for coagulation and flocculation except as they add to such other processes as either air flotation or sedimentation. Somewhat dated (1972) costs are given in Stanford Research Report No. 80 (13) for petrochemical wastewater treatment.

Application to High P/O/T Gasification--

Processes like Synthane and Lurgi that produce high phenols, oils, and tars, (p/o/t) may have as-formed alkalinities of 7,000 to 16,000 mg/l, expressed as $CaCO_3$ (3, 4). Even so, calcium and magnesium hardness may be low (15 to 20 mg/l) so that most of the alkalinity is associated with ammonium compounds. Stripping these waters for H_2S and NH_3 removal, with or without phenol extraction, also strips CO_2 and reduces the high original alkalinity to low residual levels. Coagulation and flocculation of these stripped waters to reduce suspended solids and oils may require alum treatment or addition of carbonate alkalinity so that lime treatment can be effective.

Limited results of C&F treatment of gasifier wastewaters were found in the literature (4) and some coke-liquor (5, 7) and refinery data (6, 8, 9) are also available. Synthane PDU

TABLE 8-23. RELATIVE COSTS OF COMMON pH ADJUSTMENT REAGENTS* (1977)

Reagent	Chemical Formula	Relative Cost
Alkaline		
Caustic Soda	NaOH	8.50
Soda Ash	Na_2CO_3	4.10
High Calcium Hydrated Lime	$Ca(OH)_2$	1.37
Dolomitic Hydrated Lime	$Ca(OH)_2 \cdot MgO$	1.06
High Calcium Quicklime	CaO	1.00
Dolomitic Quicklime	CaO.MgO	0.85
High Calcium Limestone	$CaCO_3$	1.21
Dolomitic Limestone	$CaCO_3.MgCO_3$	1.05
Acidic		
Sulfuric Acid	H_2SO_4	1.00
Nitric Acid	HNO_3	4.36
Hydrochloric Acid	HCl	2.30
Sulfur Dioxide	SO_2	1.47
Carbon Dioxide	CO_2	-

*Item 12 on reference list, page III, D-5 wastewater studies (4) had total influent tar, oils, and grease of 1,100 mg/l. Alum dosage of 100 to 150 mg/l, coupled with pH adjustment, reduced this to a total of about 600 mg/l, all soluble. Thus, although only about 50 percent of total tars, etc. were removed, no suspended tars, etc. remained. See the case history of the Synthane Gasifier.

One of the coke liquor references (5) is an interesting recent development of a patented process by Bethlehem Steel. Figure 8-24 shows that coke plant weak ammonia liquor (WAL) is treated in a still for ammonia recovery prior to combining with other wastewaters for biological oxidation (biox) treatment. Figure 8-25 shows the conventional WAL distillation flowsheet. Bethlehem's new process, shown in Figure 8-26, involves liming the WAL and clarifying it before ammonia stripping. To apply this process to coal conversion wastewaters, two-stage sour water stripping would be required with liming and clarification between the first, or H_2S, stage and the second, or NH_3 stage. More details are given in the case history of the Bethlehem process. This modification should be compatible with U.S. Steel's "PHOSAM-W" process for stripping and ammonia recovery but will have the added costs of a separate H_2S still. Bethlehem claims that oil and tar removal is accomplished without flotation or filtering equipment with oil, etc. coming down with the lime precipitate. One potential weakness of this process is suggested (7) in a discussion of biox treatment of WAL where the authors say that calcium thiocyanate formed during lime distillation is about "four times more difficult to (bio) oxidize" than ammonium thiocyanate. Pilot or bench tests are needed to assess the effect of lime distillation on downstream treatability.

Beychok (6), Ford (8) and Lin (9) give summaries of refinery sour water oil C&F experience. They did not distinguish soluble from "free" oil. Beychok reports that 30 to 150 mg/l alum provided 75

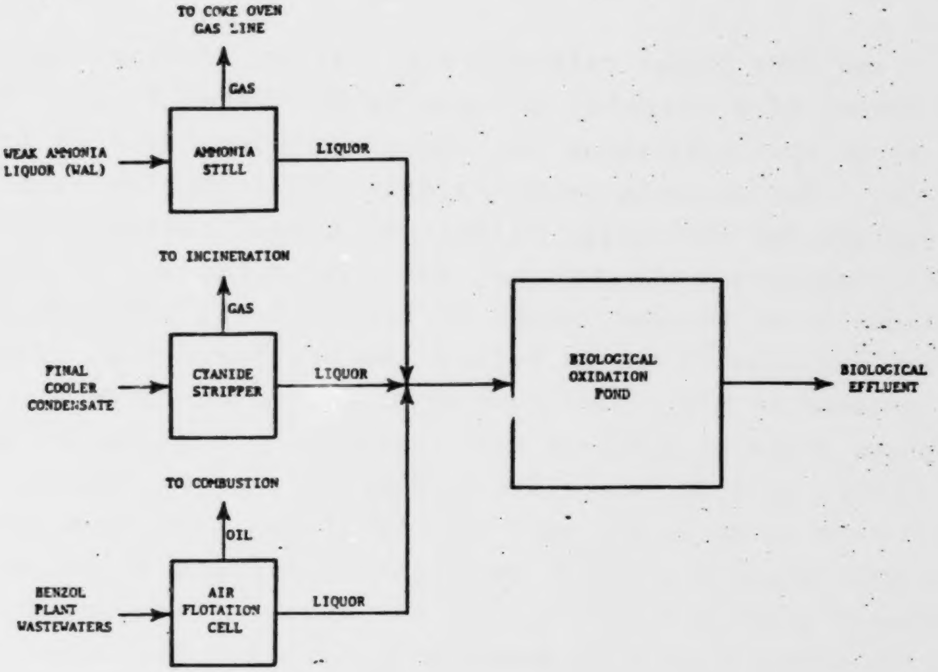
TO COKE OVEN GAS LINE
GAS
WEAK AMMONIA LIQUOR (WAL)
AMMONIA STILL
LIQUOR
TO INCINERATION
GAS
FINAL COOLER CONDENSATE
CYANIDE STRIPPER
LIQUOR
TO COMBUSTION
OIL
BENZOL PLANT WASTEWATERS
AIR FLOTATION CELL
LIQUOR
BIOLOGICAL OXIDATION POND
BIOLOGICAL EFFLUENT

Figure 8-24. Bethlehem multitreatment scheme.*

*From Item 5 in reference list

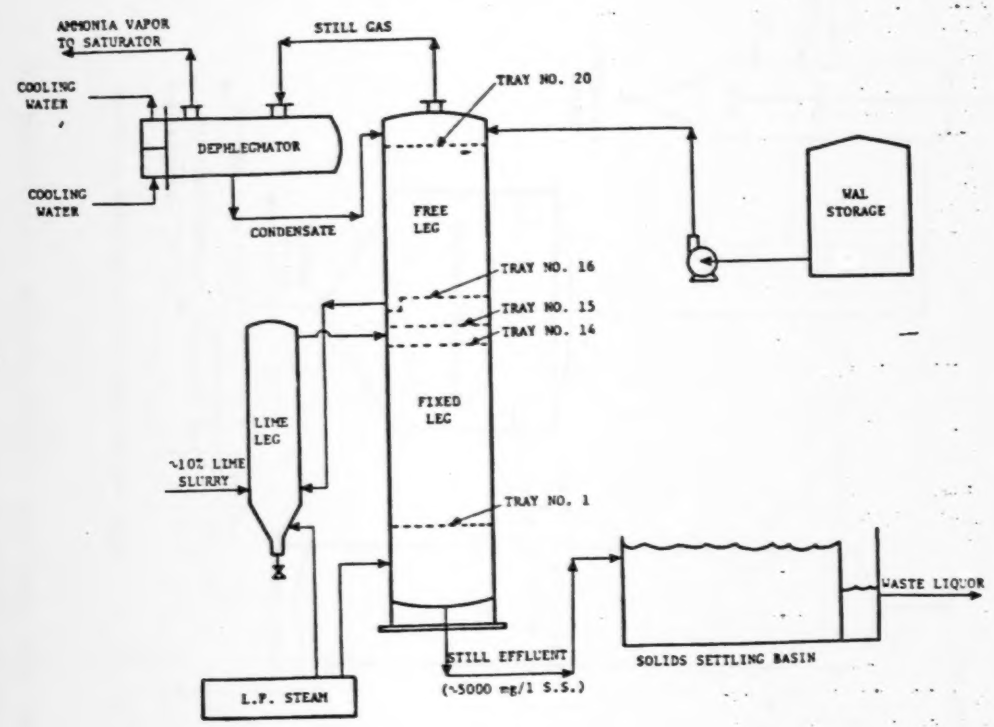
AMMONIA VAPOR TO SATURATOR
STILL GAS
COOLING WATER
DEPHLEGMATOR
COOLING WATER
CONDENSATE
TRAY NO. 20
FREE LEG
WAL STORAGE
TRAY NO. 16
TRAY NO. 15
TRAY NO. 14
LIME LEG
FIXED LEG
~10% LIME SLURRY
TRAY NO. 1
WASTE LIQUOR
STILL EFFLUENT
(~5000 mg/l S.S.)
SOLIDS SETTLING BASIN
L.P. STEAM

Figure 8-25. Conventional flowsheet for ammonia distillation.*

From Item 5 in reference list

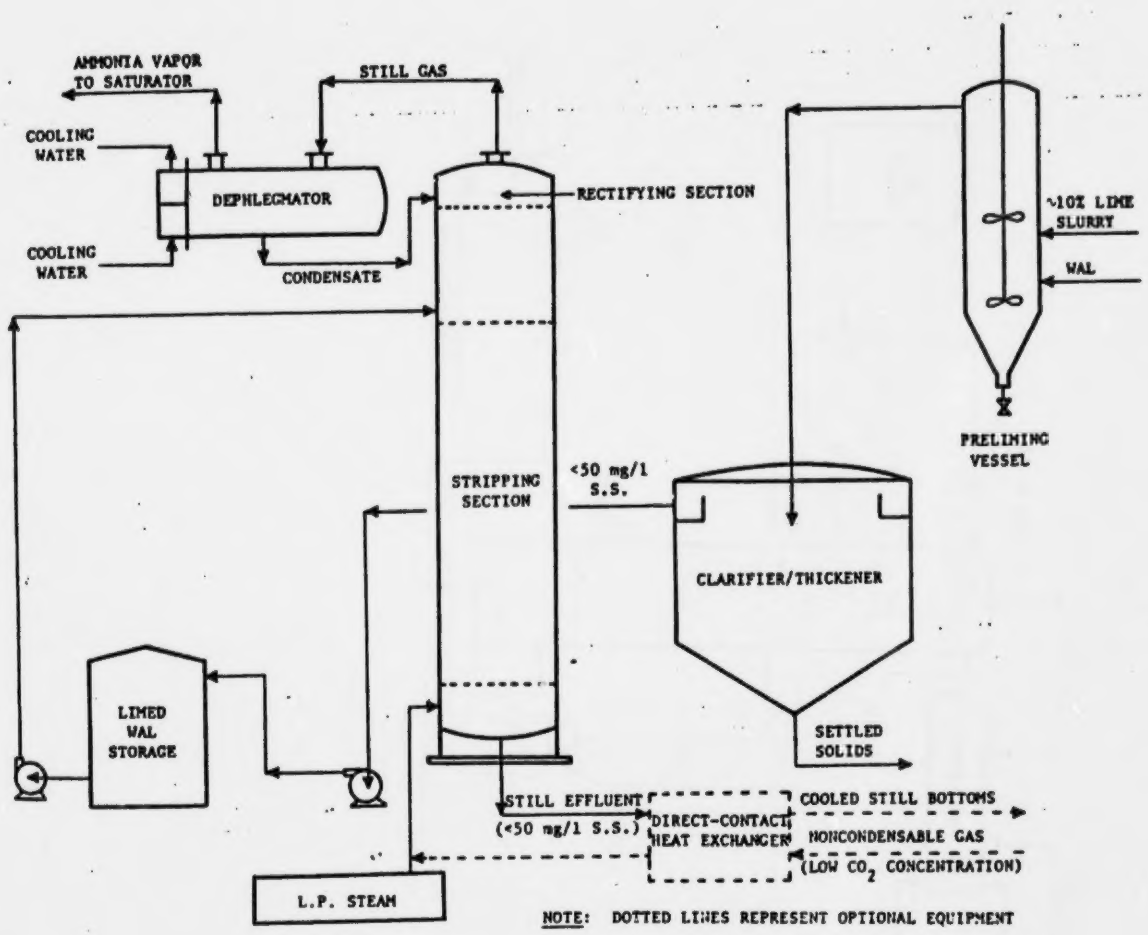
AMMONIA VAPOR TO SATURATOR
STILL GAS
COOLING WATER
DEPHLEGMATOR
COOLING WATER
CONDENSATE
RECTIFYING SECTION
~10% LIME SLURRY
WAL
PRELIMING VESSEL
STRIPPING SECTION
<50 mg/l S.S.
CLARIFIER/THICKENER
LIMED WAL STORAGE
SETTLED SOLIDS
STILL EFFLUENT
(<50 mg/l S.S.)
DIRECT-CONTACT HEAT EXCHANGER
COOLED STILL BOTTOMS
NONCONDENSABLE GAS
(LOW CO2 CONCENTRATION)
L.P. STEAM
NOTE: DOTTED LINES REPRESENT OPTIONAL EQUIPMENT

Figure 8-26. Flowsheet for ammonia distillation according to redesigned process.*

From Item 5 in reference list

to 85 percent oil removal and 55 to 70 percent suspended solids removal. Beychok states that the C&F process can produce refinery waters with no more than 20 to 30 mg/l oil. Ford reports that chemicals (alum, most commonly) at 100 to 130 mg/l reduced oil in refinery waters 71 to 93 percent when used with dissolved air flotation (DAF). Lin and Lawson give 50 to 90 percent removal of refinery oil at 60 mg/l oil influent for C&F with DAF.

Application to Gasification Producing No P/O/T--
The Koppers-Totzek process is a commercial gasification process producing no p/o/t with a reported (3) wastewater alkalinity of 700 mg/l equivalent $CaCO_3$ (calculated from the Ca and Mg hardness). Such waters require softening before final stripping to avoid scale formation in the strippers. Since H_2S stripping is greatly reduced at elevated pH, lime softening should be performed between first-stage H_2S and second-stage NH_3 stripping. Since CO_2 will strip with H_2S, carbonate alkalinity (soda ash or recarbonation) may have to be added following liming to provide effective softening. Because this water is low in BOD, downstream treatments are likely to be physical-chemical rather than biox.

Applications to Liquefaction--
Liquefaction processes will produce essentially CO_2-free sour water from the dissolver off-gas condensates. These sour waters thus have ammonia associated with sulfides and phenols and are low in hardness and carbonate alkalinity. TABLES 8-24 and 8-25 show analyses for unstripped H-Coal and SRC sour waters obtained, respectively, from a PDU (HRI, Trenton, N.J.) and a 50 TPD pilot plant (PAMCO, Fort Lewis, WA). Carbonate was not reported for either water, calcium was only 0.47 mg/l for SRC and magnesium was 0.13 and 0.7 for SRC and H-Coal, respectively. Stripping such waters after dephenolization should produce relatively soft water of low alkalinity and hardness with good treatability.

TABLE 8-24. ANALYSIS OF SOUR WATER FROM H-COAL PDU*

Parameter	Value (mg/l)
BOD, Total	52,700
Soluble	51,200
COD, Total	88,600
Soluble	88,000
NH_3-N	14,400
Organic N	51
Phenol	6,800
Sulfide	29,300
Oil and Grease	608
pH	9.5
Pb	0.06-2.90
Ni	0.10-1.70
Mo	0.01-0.50
Co	0.01-0.50
Cu	0.02-0.40
Cd	0.8
Fe	1.2
Al	2.5
Mg	0.7
Zn	0.45

*Item 10 in reference list

TABLE 8-25. ANALYSIS OF SOUR WATER FROM SRC I PILOT PLANT (a)

(mg/l unless noted)
Kentucky Coal Feed
Analyses by Water Purification Associates and Pittsburg & Midway

	pH=8.6	pH=8.2
Total Carbon	9,000	8,160
Total Organic Carbon	6,600	7,390
Inorganic Carbon	2,400 (b)	770 (b)
BOD (5 days)	32,500	
BOD (15 days)	34,500	
BOD (20 days)	34,500	
COD	43,600	25,000- 30,000
Phenol as C_6H_5OH	5,000	12,000
Total Kjeldahl N	8,300 (c)	15,000 (c)
Total Ammonia as N	7,900	14,000
Total Ammonia (meq/l)	465	824
Cyanide as CN	10	
Total Sulfur as S	10,500 (c)	16,200 (c)
Ca	0.47	
Mg	0.13	
Si	< 0.5	

(a) Item 11 on reference list.

(b) By difference, see Appendix 1, Reference 3.

(c) 22 analyses for N and S made between 10/5/75 and 12/9/75 were supplied by Pittsburg and Midway. Four of these analyses had extreme values and were arbitrarily eliminated. For the remaining 18 analyses the average total nitrogen was 12,600 mg/l with a standard deviation of 7,000 mg/l which is very random. The average ratio (moles NH_3/(H_2S) was 2.0 with a standard deviation of 0.17 which is quite reproducible.

Lime distillation of ammonia may not be necessary after phenol and H_2S removal but it will aid in removal of oils and tars.

A few limited undiluted sour water analyses have recently been published for SRC II, as shown in TABLE 8-26. These analyses are similar to those for SRC I and H-Coal shown in TABLES 8-24 and 8-25 in their content of ammonia, sulfur, and phenol.

Case History of the Synthane Gasifier--
A recent report of treatability studies made on the Synthane PDU waters includes C&F pretreatment data. This report (8), one of a few published coal gasification wastewater treatment studies, describes Synthane gasifier wastewaters: "(they) are generally light amber in color initially but darken on standing. They have a strong ammonia and cresol-like odor, resembling the odor of coke-oven byproduct waters. Chemical analyses of Synthane waters....are similar to coke plant wastes..."

This study found that good consistent oil and tar removals were obtained with 100 to 150 mg/l alum. Jar tests were used for C&F evaluation. Initial concentrations of 22,000 mg/l tar, oil and grease were reduced to 1,100 mg/l by 3 to 6 hours settling at ambient temperature. Jar tests were made with alum on the 1,100 mg/l "feed" water, with and without pH adjustment. Best results were 47 percent removal (about 500 mg/l) of oil and tars, obtained with 100 to 150 mg/l alum and adjustment to pH 1.5 to 2.5 before alum addition (pH was then adjusted back to neutral before analyses were made). Effluent oil and tars (authors say "all soluble") were thus about 600 mg/l. Subsequent biotreatability studies proved this effluent was amenable to biox treatment.

TABLE 8-26. AVERAGE ANALYSES FOR SOUR WATER IN SRC II PILOT PLANT, 1977*

	JULY	AUGUST	SEPTEMBER
	(weight percent)		
Nitrogen	3.56	4.00	4.33
Sulfur	2.55	2.51	2.97
Phenol	0.52	0.85	0.50

*Item 11 on reference list.

Case History of Bethlehem's "Improved Process" for Weak Ammonia Liquor (WAL)--

In a recent article (5) Rudzki, Burcaw and Horst describe a coke liquor pilot plant ammonia still process development program. Conventional WAL treatment for coke wastewaters uses lime addition into a "lime leg" still fed from the bottom of the upper or "free leg" still. Saturated (scaling) slurry from the lime leg feeds the top of the "fixed leg" (bottom part of the ammonia still). Plugging is common, design loadings are low (3 to 5 gpm per square foot) and steam rates are high (2 to 3 lb/gal WAL). Bethlehem's process introduces a preliming vessel and a clarifier into this process between the treated WAL storage vessel and the ammonia still feed plant. In this proces all distillation is "fixed-leg" but the feed is no longer a scaling slurry. Figures 8-25 and 8-26 are schematics of the conventional and the Bethlehem processes, respectively. The authors claim mechanical design improvements, including oil and tar removal without flotation or filtration, lower capital and maintenance costs and reduced steam consumption (0.8 to 1.1 lb/gal WAL) as advantages of the redesigned process.

TABLE 8-27 summarizes (1977) operating cost estimates for 0.2 MGD WAL plants and shows that the Bethlehem process saves $340,000/yr in operating costs. Although actual data are not given, lime costs in TABLE 8-27 indicate 32 percent more lime is used in the new process than in the old, increasing the massive lime dosage from 10,000 to 13,000 mg/l CaO (assuming 365 days operation). The clarifier/thickener produces effluent feed for the still with only 50 mg/l residual suspended solids and produces an under-flow sludge with 30 percent solids. Tar content in the sludge was 0.1 percent (dry basis) if WAL was stored prior to use as feed, but went as high as 10 percent when flushing liquor was used directly as a feed. Tar content of 0.1 percent sludge, dry basis, is equal to 26 mg/l removed from the 200,000 gpd assuming all CaO reacted

TABLE 8-27. ESTIMATED ECONOMIC ADVANTAGES OF THE BETHLEHEM AMMONIA REMOVAL SYSTEM*

Plant Capacity-757 cubic meters per day WAL
(200,000 gpd WAL)

	Operating Costs, $ per year		
Utility	Conventional Process	Bethlehem Process	Savings
Steam at $7.72/ 1,000 kg ($3.50/ 1,000 lb)	639,000	286,000	353,000
Lime at $38.55/ metric ton ($35/ton) 85% Available CaO	125,000	165,000	-40,000
Cooling Water at 8¢/10 cu meter (3¢/1,000 gal)	12,600	12,600	-
Mechanical Cleaning	30,000	Not Required	30,000
TOTAL	806,600 (2.92)**	463,600 (1.68)**	343,000 (1.24)**

* Item 5 in reference list.

**$ per cubic meter WAL treated
(to obtain $ per 1000 gallons multiply by 3.8)

to $CaCO_3$. Of course, 10 percent tar content implies 2,600 mg/l removed but the massive lime dosage employed apparently was capable of handling it. Note that tars and oils not skimmed in upstream steps are lost to lime sludge and may be difficult to recover.

Case History of an Integrated Refinery/Petrochemical Installation--

In a study of the role of industrial pretreatment for organics removal from industrial wastewaters (14), Eller and Gloyna followed bench scale studies with pilot plant work. Two streams were selected for study because of their pH and oil and suspended solids (ss) content and variability. The authors noted that after blending the two streams co-precipitated some soluble organics when the pH was increased. Influent pH varied from 2 to 10, averaging 6.0, whereas oil and ss varied from 200 to 700 mg/l. Bench scale flotation was unsuccessful and gravity operation after neutralization was marginal. Above pH 8.0 a settlable floc formed. A comparison in the pilot plant of lime and caustic for pH control showed that lime was more effective in forming settlable flocs. Lime doses of 250 to 900 mg/l gave pH 8 to 11. Optimum settling occurred with 350 mg/l at 8.5. Downstream biox reduced the pH back to 7.0 and produced a pH-stable effluent so that soda ash addition was not necessary.

The authors varied clarifier overflow rates from 400 to 1,100 gpd per square foot but found 600 gpd per square foot to be the maximum rate at which excess oil and sludge discharges could be avoided when the sum of oil and ss excursions in the influent rose to over 700 mg/l. To prevent discharges during influent excursions it was necessary to add powdered limestone as a weighting agent.

With 600 gpd per square foot overflow rate and limestone weighting as needed, the sum of oil and SS in the effluent was consistently less than 50 mg/l even with influent excursions over 700 mg/l, representing 75 to 93% removal. COD (50% probability) was about 4200 mg/l in the influent and about 2,500 mg/l in the effluent, representing about 40% removal.

References--

1. EPA 625/1-75-003a "Process Design Manual for Suspended Solids Removal," January 1975. Chapters 4, 5. 873*.

2. AIChE Today Series "Advanced Wastewater Treatment," W. W. Eckenfelder, Jr., P. A. Krenkel and C. E. Adams, Jr., AWARE, Inc., 1974. Pages D-2, D-3.

3. EPA 600/7-77-065, "Water Conservation and Pollution Control in Coal Conversion Processes." Water Purification Associates. 480*

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13. "Treatment of Petrochemical Wastewaters," E. D. Oliver, Stanford Research Institute Process Economics Program Report No. 80, September 1972. 804*

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Flotation

Flotation (1) is used to separate small amounts of fine particles or droplets from large quantities of a liquid stream or wastewater and to thicken some sludges. Except for sludge thickening this description is based primarily on petroleum refinery experience. In flotation, very small gas bubbles are released beneath the liquid surface. Rising bubbles adhere to the small particles and float them to the surface for collection. Chemical flocculants and polymers may be added to enhance collection, although in oil flotation it is desirable to avoid inorganic flocculants so the floated oil can be directly recovered or burned.

In practice, alum (2,3,4) or other coagulants are added at 25 to 200 mg/l together with < 1 to 40 mg/l polymer (2,3,4,5,6). Other appropriate design parameters are air-to-water or air-to-solids ratios of 0.01 to 0.1 lb/lb, hydraulic loadings of 500 to 4,000 gpd per square foot (2,3,4), air pressure of 25 to 70 psig (1,2,3), float detention time of 20 to 60 minutes (1,2,3), recycle ratio of 20 to 50 percent (1,2,5) and, for sludge thickening, a solids loading of 1.3 to 7.7 lb/hr per square foot (4,7).

There are three flotation systems in use: dissolved air flotation (DAF) with and without recycle and induced air flotation (IAF). DAF with recycle is widely used for both particle separation and sludge thickening (9). IAF is preferred by Chevron (6) for particle separation.

Figure 8-27 is a flow schematic of DAF with recycle used for sludge thickening.

The cleaned effluent is split into "unit effluent" and recycle streams and the recycle stream is pumped into an air dissolution

Courtesy Komline-Sanderson

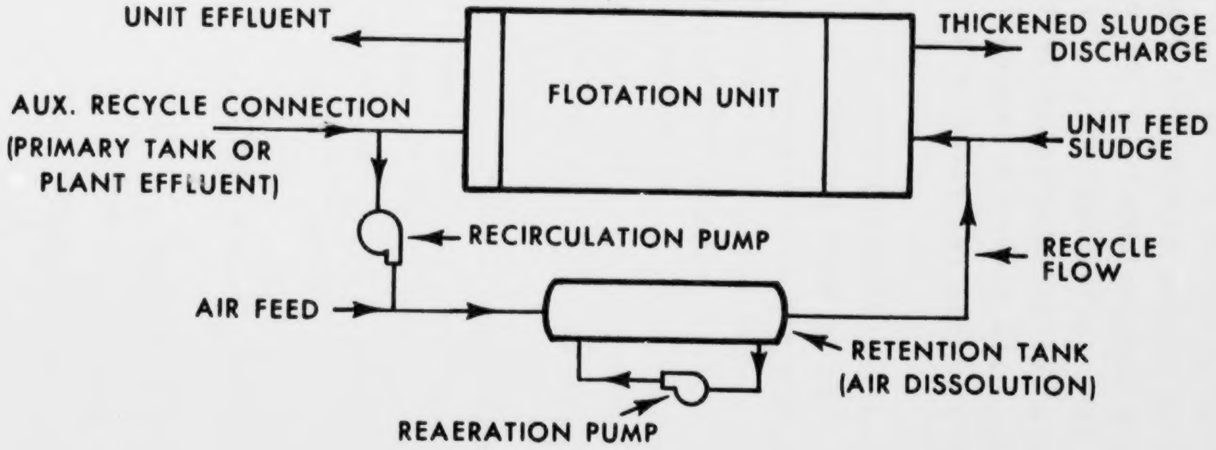
UNIT EFFLUENT
AUX. RECYCLE CONNECTION
(PRIMARY TANK OR PLANT EFFLUENT)
FLOTATION UNIT
THICKENED SLUDGE DISCHARGE
UNIT FEED SLUDGE
RECIRCULATION PUMP
RECYCLE FLOW
AIR FEED
RETENTION TANK (AIR DISSOLUTION)
REAERATION PUMP

Figure 8-27. Dissolved air flotation system.*

*From Item 7 in reference list

tank under pressure before joining the unit feed sludge stream. Pressure release in the unit allows dissolved air to leave the solution and float the solids.

Although they differ physically and in capital costs DAF and IAF systems rely on the same principles for operation. In comparison to DAF, however, IAF has the disadvantages of large minimum bubble size of about 500 to 1,000 micrometers versus 50 to 100 micrometers for DAF (1,8) and reshearing of the oil droplets and flocs by the rotor. Although IAF capital cost is probably about 50 percent of DAF capital cost, DAF with recycle is considered the system of choice for oily waters and sludge thickening since it produces fine bubbles and thus less float volume (1) and avoids emulsification of oils in dirty wastewater.

Oil Removal--

Oliver (2, chapter 5) provides an equipment list and costs for DAF for oil removal. Major equipment sizes and utilities for a 2 MGD capacity DAF operating on API separator effluent are given in TABLE 8-28. Expected efficiencies are 85 to 95 percent oil removal and 70 to 75 percent solids removal. Operating conditions do not favor extensive oxidation of BOD and COD during flotation. Power requirements are estimated at less than one hp/MGD. The design is based on residence times of 15 and 20 minutes for flocculation and flotation, respectively. Capital and operating costs were based on twin flocculation and flotation chambers, respectively 14 feet x 38 feet and 14 feet x 60 feet, each with a single 4 foot exit section. Depth was 7 feet with 2 feet freeboard. Walls were 8 inch reinforced concrete and floors were 6 inch.

Chevron (6) has extensive experience with IAF for both oil separation (secondary de-oiling) and effluent polishing (solids

TABLE 8-28. AIR FLOTATION.
MAJOR PROCESS EQUIPMENT AND UTILITIES SUMMARY*

Plant Capacity: 2 Million U.S. Gal/Day
Wastewater at 0.5 Stream Factor
(4MGD Design Flow)

Major Process Equipment

Name	Size	Material of Construction
Special Equipment		
Flocculation chambers	4 MGD	Concrete
Flotation chambers	4 MGD	Concrete
Scum skimmers (2)	14 x 60 ft	Carbon steel
Precoat filter	25 sq ft	Carbon steel
	Volume (gal)	
Tanks		
Alum	2,500	Rubber lined
Vessels		
Pressure retention vessel	3,600	Carbon steel
	Size	
Miscellaneous Equipment		
Flash mixers (2)	2 bhp	Carbon steel
Paddle mixers (6)	1 bhp	Carbon steel
Screw conveyors (2)	15 ft	Carbon steel

Pumps

Feed - 4, including 2 operating, 2 spares, 22 operating bhp.

Process - 4, including 2 operating, 2 spares, 78 operating bhp.

Utilities Summary**

	Battery Limits Total	Feed Section	Process Section
Electricity (kw)	100	26	74

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* Item 2 in reference list

** No peaks in utilities demands; generating facilities are sized for average consumptions removal). At one refinery, oil reduction experience on API separator effluent was as follows:

Flow, GPM	Avg. Oil Content, mg/l	
	In	Out*
3,000-4,000 Actual	410	27
3,000 Design	100	20

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*Polyelectrolyte dosage was 5 to 15 mg/l.

Thus, 80 percent removal of secondary oil was the design basis and over 90 percent was achieved.

Sludge Thickening--

Sludge thickening by flotation is based primarily on municipal DAF studies reported in an EPA document (7). In operation, unit effluent is aerated and pressurized and then a controlled flow is metered into a mixing chamber where it joins incoming sludge feed and any polyelectrolytes or other chemicals. Sludge is floated as a blanket 8 to 24 inches thick where it thickens by emergence and drainage. Thickened sludge is removed by a skimming device. Bottom sludge collectors are provided to remove grit and heavy materials.

Air pressure depends on the unit design but is optimized in the range of about 40 to 70 psig, by observing that increases in pressure result in greater separation and solids concentration up to the pressure where further increases cause floc destruction. Recycle ratio is optimized at about 40 percent. Increasing the recycle requires larger equipment, increases capital and operating costs, and dilutes the influent sludge, but dissolves more air and thus increases the air-to-solids ratio. Float solids increase, and effluent SS decrease, as float residence time increases up to, but not over, 3 hours. Finally the air-to-solids ratio influences the float solids obtained depending on the sludge quality. Figure 8-28 shows the float solids concentration versus air-to-solids ratio for several sludges. Note that a sludge viscosity index (SVI) below 100 generally indicates that the sludge will flocculate and settle well. Reference 7 also gives operational results, details for integrating a DAF sludge thickener into the wastewater treatment plant, equipment lists, a design example and generalized costs.

Both API and DAF/IAF skimmings and sludges contain recoverable oils but treatment is required to separate them. Chapter 8 of reference 10, the API Manual on Disposal of Refinery Wastes, treats oil recovery from flotation skimmings.

References--

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4. Beisinger, M.G., Vining, T.S., and Shell, G.L., "Industrial Experience With Dissolved Air Flotation," 29th Purdue Industrial Waste Conference, 1974. 886*

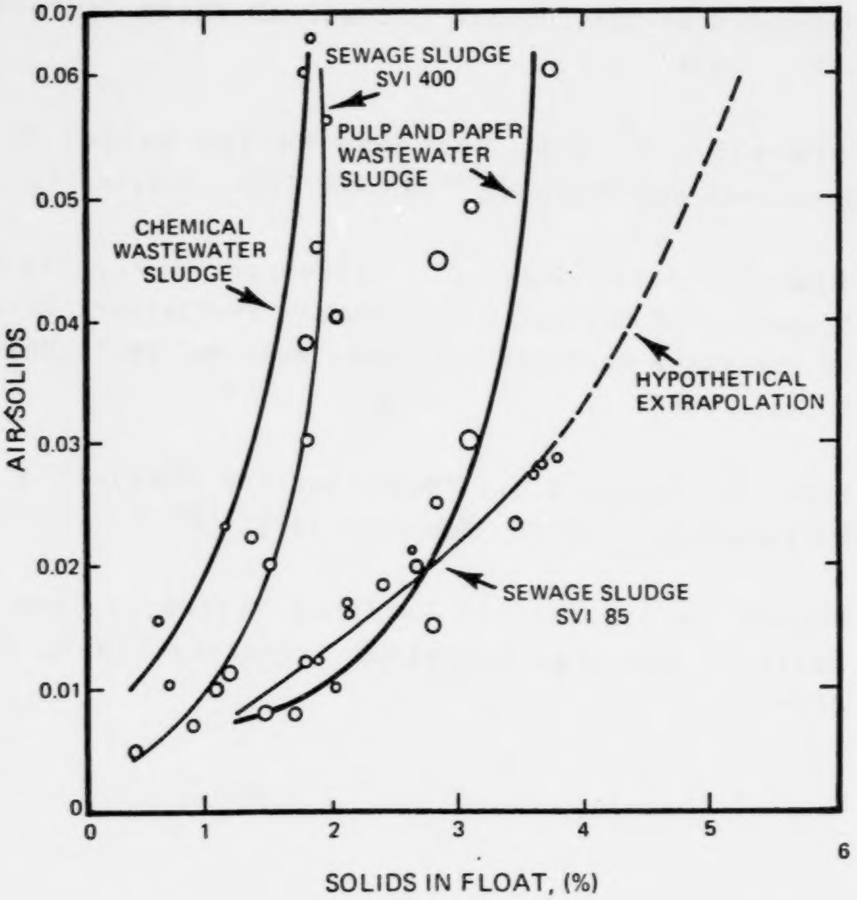

Figure 8-28. Influence of air-to-solids ratio on float solids content.*

*From Item 7 in reference list

5. Luthy, R.G., Selleck, R.E., and Galloway, T.R., "Removal of Emulsified Oil with Organic Coagulants and Dissolved Air Flotation", JWPCF, February 1978. 885*

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*Pullman Kellogg Reference File number

Biological Oxidation

Biological oxidation, the bacterial destruction of organic matter in wastewater streams under aerobic conditions, is the most important and widely used of the wastewater treatment processes. Many variations of the basic process have been developed for treatment of specific types of contaminants in waters, for increasing the efficiency of removal of organic contaminants and for decreasing costs. Most of the variations fall into the two general categories of fixed film systems, in which bacteria grow on a solid inert substrate, and suspended growth systems, in which the bacterial mass is flocculant and suspended in the reactor liquid. Fixed film systems include trickling filters, rotating biological contactors and fluid bed reactors. Suspended growth systems include activated sludge processes and aerobic sludge digestion. Rotating biological contactors and activated carbon enhanced activated sludge processes will be discussed in detail later in this section; only conventional and pure oxygen activated sludge systems are considered here.

Industrial biological wastewater systems usually include treatment steps that precede the oxidation process. Pretreatment may include separation of light and heavy oils, flow equalization, flocculation/flotation by dissolved air flotation (DAF) or induced air flotation (IAF), solvent extraction and gas stripping.

Comparison of fixed film and suspended growth systems shows that each has certain characteristic advantages and disadvantages. For example, fixed film systems are especially useful for "roughing" operations and general first-stage biological treatment because the attached bio-films are more resistant to physical and chemical shock than are flocculant suspended growths. Further, the fixed film systems produce less excess biomass (sludge) per unit of BOD removed and the sludges are more compact. On the other hand, suspended growth systems are generally capable of reducing the effluent BOD and COD to lower levels and have higher removal efficiencies for nutrients and metals as well as for dissolved organics. TABLE 8-29 compares the two systems, with the trickling filter representing fixed film and activated sludge representing suspended growth.

Biological oxidation systems have the advantages of low cost installation and operation, the ability to effectively oxidize very dilute streams and great operational flexibility. They also have many weaknesses such as susceptibility to toxicity, physical and chemical upsets, hydraulic washout, organic load variations, and equipment failures, any of which may interfere with bacterial growth and the efficiency of BOD and COD removal. In addition, biological oxidation systems cannot achieve BOD, COD, and suspended solids levels below their residual limitations and cannot effectively remove most metals and those materials that are not readily biodegradable.

Application of Biological Oxidation to Coal Wastewaters--

The sour waters from gasification processes producing phenols, oils and tars and from liquefaction processes are expected to be similar to coke waste liquors. After stripping CO_2 completely, and stripping H_2S and NH_3 down to required bacterial nutrient levels, the sour water will be pretreated for oil recovery and flow equalization and will then be introduced into a fixed film biological oxidation process. A trickling filter is assumed for this description. Systems with phenol recovery will have influent phenol concentrations below about 600 mg/l and roughing filters should readily reduce this by 30 to 60 percent (1-399*) (4).

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*Item 1 in reference list, page 399

TABLE 8-29. BIOLOGICAL OXIDATION SYSTEM EFFICIENCIES (1)

System (2)	Waste (3)	Flow MGD	Influent BOD mg/1	Influent COD mg/1	Effluent BOD mg/1	Effluent COD mg/1	Removal BOD %	Removal COD %	Remarks	Ref. (4)
TF	M	1.8	375	-	68	-	74*	-	Low rate. Strong waste	1, 4-4
TF	M	1.4	77	-	44	-	43*	-	High rate. Weak waste	1, 4-23
AAS	M	3.9	254	-	33	-	87*	-		1, 5-3
POAS	M	4.4	84	-	10	-	88*	-	Weak waste	1, 5-20
AAS	C	-	4140	-	146	-	97	-	Pilot test. Strong liquor	3, 873
AAS	L	-	1890	3070	26	360	99	88	Pilot test. Food/mass = 0.06	4, 22
AAS	L	-	2600	4180	36	310	99	93	Pilot test. Food/mass = 0.17	4, 22
AAS	L	-	2070	3180	24	380	99	88	Pilot test. Food/mass = 0.22	4, 22
TF	I	-	3000	-	750	-	75	-	Pilot test. Candy waste on Surfpac (Dow filter packing)	5, 890
AAS	G	-	-	720	-	72	-	90	Permanganate removals of commercial dephenolated Lurgi wastewater in pilot units	6, 104

(1) All efficiencies include secondary clarification. Those marked * also include primary clarifiers at 30% removal

(2) TF = Trickling filter; AAS = Air activated sludge; POAS = Pure oxygen activated sludge

(3) M = Municipal; C = Coke oven liquor; L = Liquefaction plant; I = Industrial; G = Lurgi gasification

(4) 1, 4-4 refers to Item 1 in the reference list, page 4-4

Units processing undephenolized liquor containing perhaps 6,000 mg/l of phenol will be oxygen transfer limited (1-401) for economic height to width ratios and must be designed and operated with forced ventilation. Although there is some concern that roughing filters will have to use effluent recycle for temperature control (1-399) rather than dilution and seeding, it seems possible that the large equalization basins anticipated (2-880) (3-35) will provide all the cooling needed.

Activated Sludge--

Activated sludge processes should be able to handle the same waters from coal conversion processes with or without phenol extraction, but large equalization tanks will be required for reliable operation (1-363) (3-35). In the activated sludge process the influent water stream mixes with a return stream from the secondary clarifier in the system. The return stream is part of the clarifier underflow and carries with it actively growing biomass. The mixed stream is retained sufficiently long in an aeration basin, while air is bubbled through the liquid, to allow destruction of the organic materials in the influent stream and then flows to the secondary clarifier. The clarifier overflow, devoid of biomass, is discharged for process use or further treatment as required in the plant process flow scheme. The clarifier underflow (sludge) is divided into return and waste activated sludge. Clarifier underflow is so dilute, at 1 to 3 percent solids, that no other recycle is usually required and the return flow may equal the wastewater flow. If phenol extraction is not used, it will probably be more economical to precede the activated sludge (AS) units with roughing filters or other pretreatment rather than attempt to design them to handle concentrations of 6,000 mg/lor more of phenol directly (1-396).

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*Item 1 in reference list, page 399

High Purity Oxygen Activated Sludge (HPOAS) processes are modifications of AS systems with no conceptual differences. The economic advantages may be especially attractive at coal conversion plants using onsite oxygen generation for the gasifiers. Advantages of HPOAS besides economy are plant compactness, lowered effluent residuals and complete enclosure for good air emissions control and weather protection. The advantage of good control of gaseous emissions from activated sludge processes is particularly important because of possible effects on health of polynuclear aromatic compounds in the emissions. Weather protection is important in maintaining the optimum temperature for bacterial growth, particularly in the winter months (3-35).

AS Costs--
The costs of biological oxidation depend primarily on the flow rate and concentration of the wastes and secondarily on the waste variability. Most wastewater treatment plant designs include equalization, neutralization and flotation or other pretreatment as well as secondary clarifiers after the biological oxidation stages. Costs (1976) for 3 MGD capacity plants treating stripped sour water containing 6,300 mg/l phenol were estimated at $3.2/ 1,000 gallons for air activated sludge (AAS), $3.6/ 1,000 gallons for high purity oxygen activated sludge (HPOAS) and $3.1/1,000 gallons for active trickling filter plus high purity oxygen activated sludge (ATF-HPOAS) systems (1-377,-295,-405). The HPOAS and ATF-HPOAS systems were designed as two-stage systems. The AAS was a one-stage system scaled up from coke plant weak ammonia liquor data. The costs included 7-day retention time equalization tanks for all plants. Sludge age or solids retention times were 11 days, 3 2/3 days and 3 2/3 days for AAS, HPOAS, and ATF-HPOAS systems, respectively. Comparable, i.e., one-stage air-activated sludge, plants for several coke weak ammonia liquor wastes had sludge ages from 1 to 4.8 days (1-367).

Case Histories of Biological Oxidation--

Lurgi Wastewater--R. Cooke and P. W. Graham (6) performed pilot waste treatment studies with biological and lime treatment on stripped dephenolated sour water from a Lurgi high pressure process in the early-to-mid 1960's. Conventional one and two stage air activated sludge (AAS) processes were used and removed up to 94 percent of the phenolics and 99 percent of the thiocyanates.

The sour water source was the spent liquor outlet of the ammonia still from the Dorsten, Germany Lurgi plant. Five batch deliveries of about 1,000 gallons each were made by tanker, with half delivered to each of two labs. Unfortunately, up to 7 days' storage may have occurred prior to dephenolization and ammonia recovery. The first delivery showed visible changes between delivery of half the load (500 gallons) in London and the other half subsequently in Nottingham. The liquor became darker in color and more difficult to treat. A "significant" reduction in phenol occurred. Subsequent deliveries were made in a two-compartment tanker.

Ten small samples were taken over a period of a month to determine sample variability. There was an unfortunate delay of 4 to 7 weeks between sampling and analysis for these samples.

The dephenolized liquor contained only 5 to 10 percent of the original phenol concentrations. Dephenolization via the Phenosolvan process preferentially removed monohydric phenols, leaving catechol and resorcinol as most of the remaining "phenol" content to be removed or treated. TABLE 8-30 summarizes the analyses of the feed liquor by batch and includes data on small samples taken to show feed variation. Total phenols were approximately 10 times monohydric phenols. Biotreatability was adequately shown in that two different bench scale biological oxidation air activated sludge units were successfully operated

TABLE 8-30. ANALYSES OF SAMPLES OF DEPHENOLATED LURGI LIQUOR (a)

Batch	1		2		3		4		5		Small samples	
	Unshifted		Unshifted		Shifted		Mixed		Shifted		Shifted	
Laboratory (b)	G.C.	N.C.B.	G.C.	N.C.B.	G.C.	N.C.B.	G.C.	N.C.G	G.C.	N.C.B.	G.C.	Standard deviation
Free ammonia (NH_3)	220	50	190	140	153	130	79	110	112	-	1150	148
Fixed ammonia (NH_3)	1140	1190	1190	1470	1530	1560	1050	1140	1250	-		
Monohydric phenols (C_6H_5OH)	56	40	18	29	16	24	12	34	18	-	19	3
Total phenols (C_6H_5OH)	390	263	284	285	228	259	258	284	277	-	209	32
Thiocyanate (CNS)	161	164	222	227	86	97	181	182	67	-	74	5
Thiosulphate (S_2O_3)	105	-	23	-	95	-	53	-	79	-	37	14
Sulphide (S)	-	-	Nil	-	4	-	2	-	1	-	-	-
Organic carbon (C)	676	-	555	-	323	-	482	-	353	-	-	-
Chloride (Cl)	2400	2300	2340	2360	2700	2690	2210	2090	2230	-	2110	95
Permanganate value (4 hr)	710	607	686	603	503	458	558	512	491	-	374	30
pH value	7.1	6.9	8.0	8.2	8.1	8.1	7.8	7.9	7.2	-	8.1	0.3
Bacterial count (thousands of organisms per ml)	620	-	450	Sterile	Sterile	0.65	110	Sterile	Sterile	-	-	-

Concentrations in ppm

(a)From Item 6 in reference list

(b)G.C. = Gas Council; N.C.B. = National Coal Board on Lurgi wastes. Figures 8-29(a) and 8-29(b) show flow schematics, respectively, for the Gas Council and National Coal Board units. Results were quite similar despite differences in the units. TABLES 8-31(a) and 8-31(b) summarize data for biotreatment performed, respectively, before and after ammonia removal. Both units achieved approximately 99 percent removal of thiocyanate and about 94 percent total phenolics removal. Lime treatment was applied by boiling the effluent with lime. Biotreatability was adequate when biological oxidation preceded and also when it followed lime treatment. In general pre-limed effluents were easier to treat biologically than un-limed effluents. Many other studies were performed as a part of this work, including BOD removal, nutrient requirements and toxicity and bacterial tests. The authors concluded that biological treatment can be performed with no great difficulty.

Synthane Wastewaters--The Pittsburgh Energy Research Center of DOE is studying Synthane wastewater treatability. In a recent report (7) the wastewaters from Synthane, a producer of p/o/t, were described as similar in appearance and analysis to coke plant wastes, being dark colored and odorous. Laboratory pretreatment was used to prepare Synthane PDU wastewaters from Montana Rosebud coal for biological oxidation studies. Pretreatment included settling, pH adjustment, alum addition, and air stripping to about 500 mg/l residual NH_3.

Laboratory activated sludge biological oxidation units were seeded with coke plant sludge to start them. Studies were based on total organic carbon (TOC) but COD and phenol were also measured. Food to microorganism ratio (F/M) was varied by varying feed dilution, an undesirable technique because inhibitory substances may affect treatability variably, depending on the degree of dilution. Dilution was necessary because the supplies

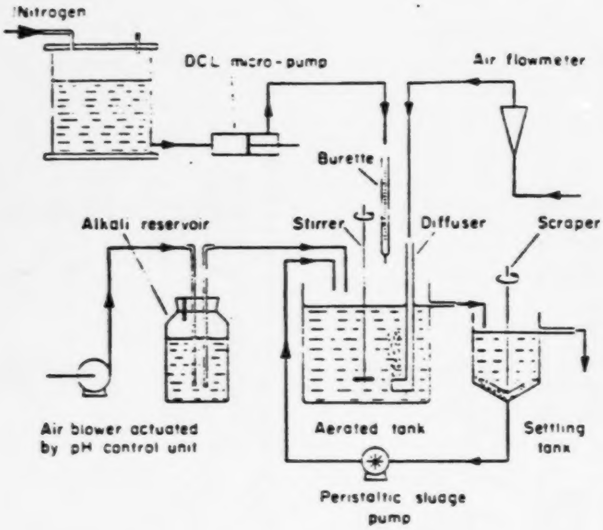
Nitrogen
DCL micro-pump
Air flowmeter
Burette
Alkali reservoir
Stirrer
Diffuser
Scraper
Air blower actuated by pH control unit
Aerated tank
Settling tank
Peristaltic sludge pump

Figure 8-29(a). Flow diagram of Gas Council apparatus.*

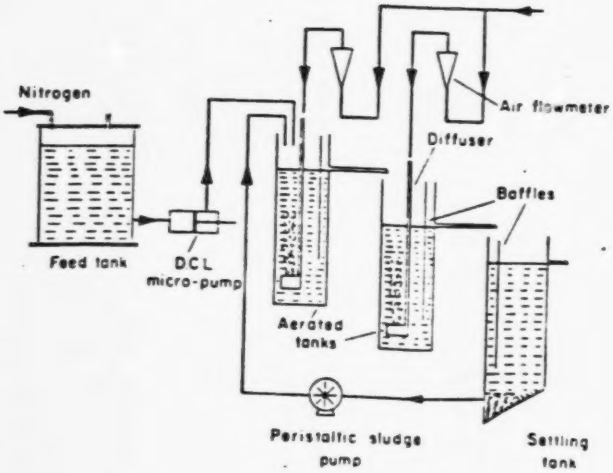
Nitrogen
Air flowmeter
Diffuser
Baffles
Feed tank
DCL micro-pump
Aerated tanks
Peristaltic sludge pump
Settling tank

Figure 8-29(b). Flow diagram of National Coal Board apparatus.*

*From Item 6 in reference list

TABLE 8-31. AVERAGE RESULTS OF TREATMENT*

(a). Biological Treatment followed by ammonia removal

	Untreated liquor	Biologically-treated effluent		Post-limed effluent	
	(ppm)	(ppm)	Removal (%)	(ppm)	Removal (%)
Total ammonia (NH_3)	1410	1490	--	90	--
Monohydric phenols (C_6H_5OH)	27	3	89	3	89
Total phenols (C_6H_5OH)	281	63	78	18	94
Thiocyanate (CNS)	154	2	99	1	99
Chloride (Cl)	2370	2440	--	2480	--
Permanganate value (4 hr)	570	127	78	50	91
pH value	7.7	7.4	--	7.3	--

(b). Ammonia removal followed by biological treatment

	Untreated liquor	Pre-limed liquor		Biologically treated pre-limed effluent	
	(ppm)	(ppm)	Removal (%)	(ppm)	Removal (%)
Total ammonia (NH_5)	1410	40	--	101	--
Monohydric phenols (C_6H_5OH)	27	16	41	2	93
Total phenols (C_6H_5OH)	281	66	77	23	92
Thiocyanate (CNS)	154	147	5	2	99
Chloride (Cl)	2370	2320	--	2430	--
Permanganate value (4 hr)	570	293	49	57	90
pH value	7.7	8.0	--	7.7	--

*From Item 6 in reference list

of wastewater from the Synthane PDU were limited. Nutrients were supplied by phosphate-supplemented settled primary municipal sewage. Hydraulic detention time was one day and steady state operation was usually attained in four days.

Results were presented for COD, phenol, and TOC for two F/M ratios and are tabulated below. Note the relatively high effluent residuals for COD and TOC.

SYNTHANE BIOLOGICAL TREATABILITY

	F/M	Influent mg/l	Effluent mg/l
COD	0.71	5,690	2,030
Phenol	0.71	1,205	25
TOC	0.71	1,900	850
COD	0.20	1,250	390
Phenol	0.20	175	<1
TOC	0.20	500	150

In summary, Synthane wastewaters are clearly amenable to treatment by biological oxidation, although dilution and nutrient addition via municipal sewage may have affected results slightly. High residuals from one-stage treatment may require second stage AS and/or tertiary treatment steps.

H-Coal Wastewaters-- The H-Coal liquefaction process hydrotreats coal slurry in an ebullated bed catalytic reactor at about 850° F and 3000 psig. Sour waters are produced from oxygen and residual moisture in the coal and condense and separate when the reactor effluent gases are processed for product recovery. AWARE, Inc. (8) designed an AS wastewater treatment plant for a 600 TPD pilot plant that is under construction at Cattlettsburg, Ky. Their design was based on the pilot plant design, combining various blowdowns and runoffs so the foul water constituted 22 percent of the total.

AWARE experimental work was carried out on foul water from the Trenton, N.J. PDU collected during a mild hydrogenation run appropriate for fuel oil production. Illinois No. 6 coal is inferred to be the source of wastewater from the statement that coal pile runoff was simulated by contacting Illinois No. 6. The PDU sour waters were steam stripped with reflux at first high (10.5), then low (6.5) pH. Stripped sour waters were combined with actual and simulated blowdowns and runoff and then pretreated with induced air flotation followed by neutralization for oil removal. Stripping H_2S from the sour water reduced the COD from 88,600 mg/l to 26,500 mg/l, essentially reducing the COD by 2 mg/l for each mg/l of H_2S removed. Combined wastewater was treated in 20-liter plastic AS units and studies included pretreatment for phenol removal by resin, startup/shutdown simulations and F/M ratio variations. Many other studies addressed IAF, emulsion-breaking, neutralization, stripping, oxygen transfer, sludge production, sludge handling and carbon adsorption. Representative results for H-Coal biological oxidation are tabulated on the following page.

H-COAL BIOLOGICAL TREATABILITY

	F/M	Influent mg/l	Effluent mg/l
COD	0.06	3,070	360
BOD	0.06	1,890	26
Phenol	0.06	750	0.7
COD	0.17	4,180	310
BOD	0.17	2,600	36
Phenol	0.17	1,450	0.3
COD	0.22	3,180	380
BOD	0.22	2,070	24
Phenol	0.22	760	0.7

This study shows that combined H-Coal wastewaters were biologically treatable and that the low residuals achieved imply attractive economics for water treatment for reuse or discharge.

Coke Plant Liquors-- The literature contains several reports and studies on the application of activated sludge systems to coke plant ammonia liquors (3, 9, 10,11) and at least one paper (12) describing the application of this technology to coal gasification. The latter paper by Parsons and Nolde (12) describes the historical development of biological wastewater treatment from coal gas plants to coke plants and producer gas plants. They observe that recent refinements in bio-treatment of coke plant liquor will be transferable to p/o/t-producing coal gasification processes.

The paper by an AWARE team (3) describes experiments on two coke plant wastes and compares results with EPA Coke Plant "Best Practical Control Technology Currently Available" guidelines. The authors give a summary of operating data from six full scale activated sludge plants and state that equalization and cooling should be the major design considerations. The AWARE studies found that effluent ammonia concentration was difficult to control. Low F/M ratios, below that needed for BOD removal, enhanced cyanide and phenol reduction and improved sludge characteristics.

A second AWARE paper by Carl Adams (9) presented results of activated sludge treatment studies on a synthesized waste for a coke plant flash evaporator condensate. The condensate was known to contain concentrated ammonia and phenolics. This study found phenolics could be reduced from several thousand mg/l to less than 0.5 mg/l with adequate equalization and that above F/M = 0.3 nitrification ceases.

Reference 10 is an early U.S. study of a coke liquor activated sludge pilot plant. Successful results led to design and construction of a full scale plant. An EPA report (11) details pilot studies made to achieve nitrogen removal via a two-sludge system. TABLE 8-32 summarizes some of the results of these studies.

Rotating Biological Contactors--

Rotating Biological Contactors (RBC) are a form of fixed film biological treatment process where the bio-solids grow attached to surfaces of discs that are attached to a horizontal shaft and that are slowly rotating, partly submerged, in a treatment tank. RBC's therefore have many of the characteristics of trickling filters, including resistance to sudden overloads and to slugs of toxic material.

TABLE 8-32. REPRESENTATIVE COKE LIQUOR TREATABILITY STUDIES

Ref.			F/M			Influent, mg/l			Effluent, mg/l		
			I	II*	III	I	II*	III	I	II*	III
3	Plant A	COD									
		BOD	0.24	0.61	1.05	4140	4140	4140	146	107	499
		TOC	0.24	0.61	1.05	1400	1400	1400	334	303	414
		Phenol	0.24	0.61	1.05	1.16	1.16	1.16	0.17	0.18	0.27
3	Plant B	COD	0.21	0.38		1950	1950		263	313	
		BOD	0.21	0.38		1880	1880		65	119	
		TOC	0.21	0.38		530	530		87	113	
		Phenol	0.21	0.38		430	430		0.08	0.43	
9	Evaporator Condensate	COD	0.174	0.198	0.286	7976	8233	8341	87	361	468
		BOD	0.174	0.198	0.286	6170	6368	6666	26	76	64
		TOC	0.174	0.198	0.286	2108	2260	2260	26	128	256
		Phenol	0.174	0.198	0.286	3316	3266	3266	1.0	<0.1	0.55
10	Activated Sludge	Phenol**	0.73	1.14	1.42	3450	3700	3500	0.2	0.4	0.55

* For Reference 9, Case II was the first stage of a 2 stage unit
**F/M values computed as 1.6 times the phenol loading

Figure 8-30(a) is a flow sheet for a Chevron (13) refinery wastewater treatment system incorporating RBC's. The RBC's are preceded by an equalizing tank and followed by a clarifier. Figure 8-30(b) is a schematic of this installation showing the RBC in four stages and the clarifier following the unit.

The feed water enters a tank in which the rotating surfaces or discs are partially immersed. Rotation is 1 to 2 rpm (13, 14, 16), submersion is about 40 percent and head loss is about 6 inches. Units with 200 discs 3/4 inch thick and 11 feet in diameter have a surface area of 38,000 square feet in a cell about 12 feet x 25 feet. Biological slimes grow over all the wetted surfaces and the rotation oxygenates not only the films and adherent water but also the tank contents. Staging is usually provided by internal baffles with weirs. Excess biological growth sloughs off the discs and leaves the RBC with the effluent. If suspended biological growth is avoided, the RBC sloughings will settle satisfactorily in the following clarifier. The clarifier is usually designed for 700 to 800 gallons per day per square foot (13, 16) overflow rate and an underflow solids content of 1 to 2 percent. These solids are not recycled back to the RBC in usual practice because the discs almost never shed all their growth simultaneously and recycle is not needed.

RBC Capabilities--

RBC's have several inherent advantages over other aerobic biological treatment methods, such as resistance to upsets, low energy needs, compactness, modular construction, easy staging and weather protection. The holding tank in each RBC stage provides a diluting and surge-absorbing reservoir which contributes to its stability. Most other fixed film bio-reactors do not have a comparable reservoir. Weather protection, provided by a cover, not only enhances low temperature operations and prevents ice damage but also collects aerosols and mists generated by splashing and permits easy ventilation control.

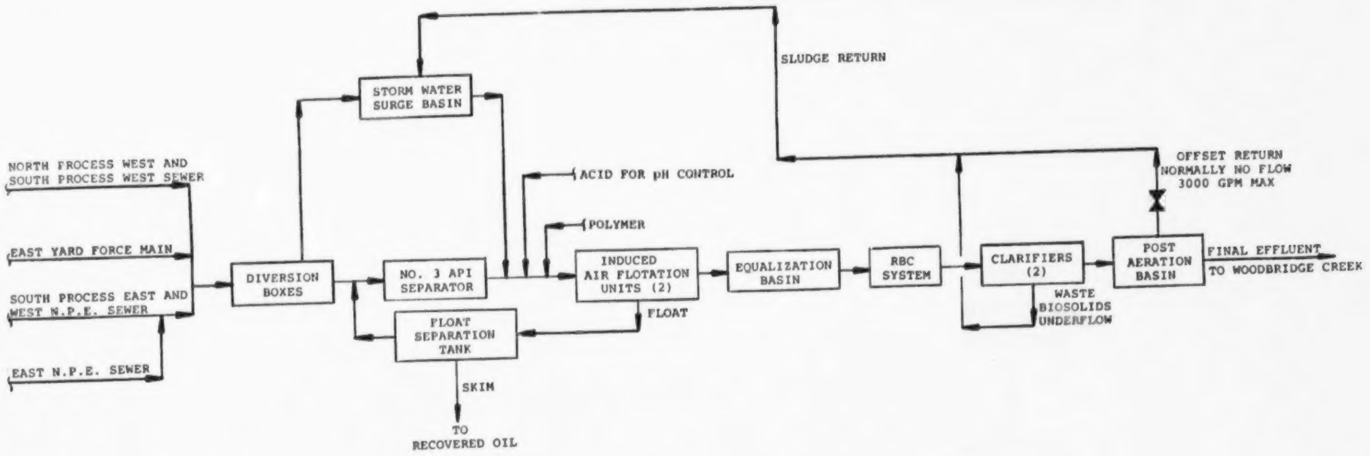
NORTH PROCESS WEST AND SOUTH PROCESS WEST SEWER
EAST YARD FORCE MAIN
SOUTH PROCESS EAST AND WEST N.P.E. SEWER
EAST N.P.E. SEWER
DIVERSION BOXES
STORM WATER SURGE BASIN
NO. 3 API SEPARATOR
FLOAT SEPARATION TANK
SKIM
TO RECOVERED OIL
ACID FOR pH CONTROL
POLYMER
INDUCED AIR FLOTATION UNITS (2)
FLOAT
EQUALIZATION BASIN
RBC SYSTEM
CLARIFIERS (2)
WASTE BIOSOLIDS UNDERFLOW
SLUDGE RETURN
OFFSET RETURN NORMALLY NO FLOW 3000 GPM MAX
POST AERATION BASIN
FINAL EFFLUENT TO WOODBRIDGE CREEK

Figure 8-30a.- Effluent treating system at Refinery B, Chevron USA.

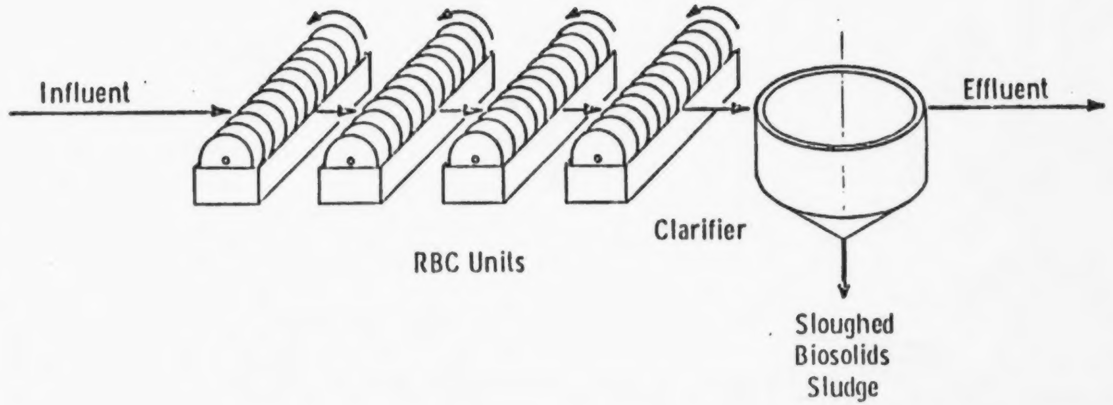
Influent
Effluent
RBC Units
Clarifier
Sloughed
Biosolids
Sludge

Figure 8-30(b). Simplified flow scheme of RBC treatment of petroleum refinery wastewater.*

*From Item 13 in reference list

RBC Applications--

RBC's have not been used commercially on U.S. coal conversion plant or coke oven wastewaters but recently they have been evaluated and installed for several U. S. refineries (13, 14, 15, 16). Applications and case histories are therefore based on refinery experience. For coal conversion wastewaters RBC's would probably be used for second stage biological oxidation, following AAS or HPOAS treatment.

Chevron pioneered the use of RBC's for refinery wastewaters. Disc area is estimated by use of McAliley's (17) method which assumes that BOD removal is proportional to BOD concentration in each stage and that the stage acts as a back-mixed reactor:

$$\begin{aligned} R &= M/A\,(BOD_{Influent} - BOD_{Effluent}) \\ &= P(BOD_{Effluent})/(K + BOD_{Effluent}) \end{aligned}$$

where R = specific BOD removal rate per unit area of medium, pounds per day per square foot

C = concentration of BOD in an RBC stage $= BOD_{Effluent}$

M = wastewater flow, million pounds per day

A = area of an RBC stage, square feet

P = a constant = maximum value of R at infinite BOD

K = a constant = BOD at which R = P/2

Figure 8-31(a) is a plot of specific removal rate versus BOD concentration for a staged Chevron RBC pilot plant. The solid curve shown was obtained from McAliley's equation. The equation can be converted into an equation of a straight line by taking the reciprocal and separating variables. Figure 8-31(b) is a plot of 1/R versus 1/C where the y-intercept and x-intercept of the best straight line through the data represent 1/P and 1/K, respectively.

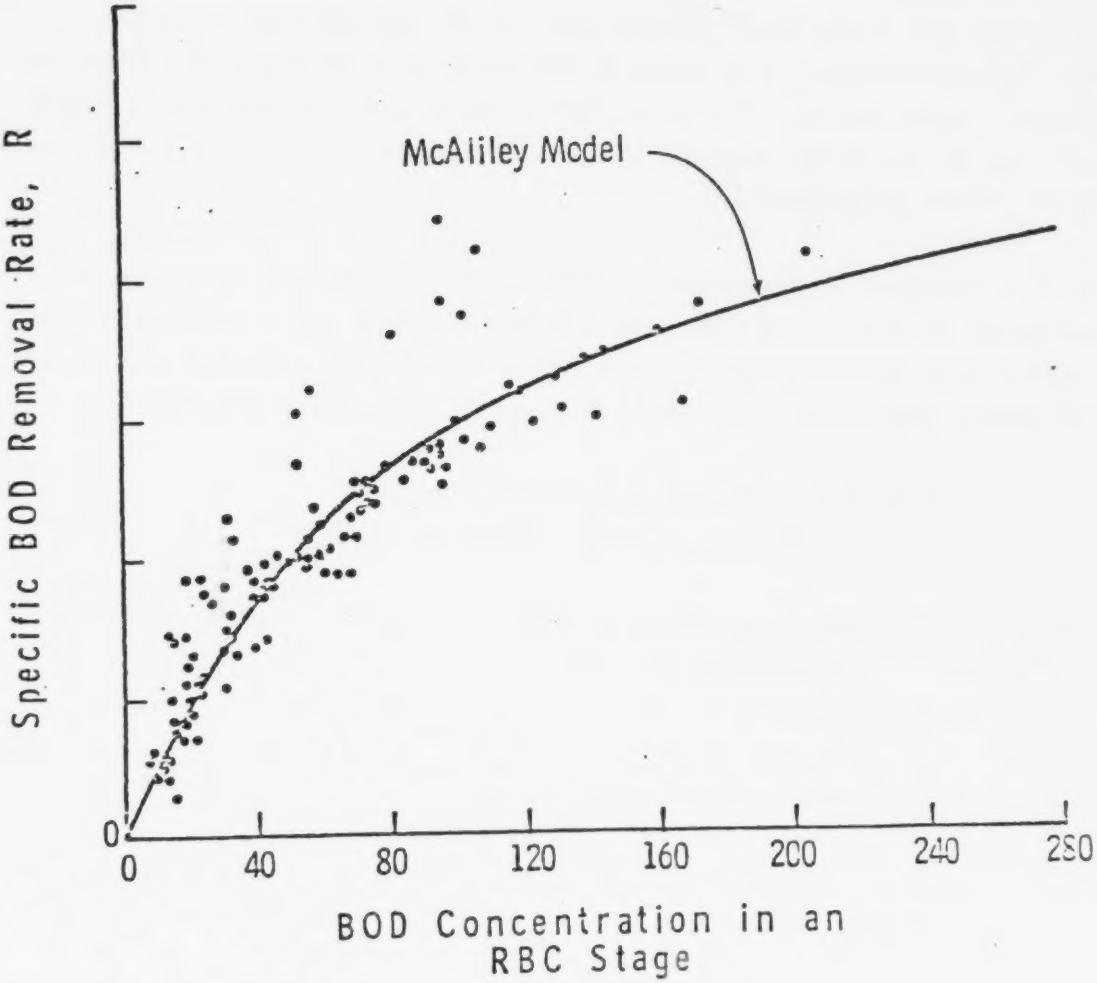

Figure 8-31(a). BOD removal rate versus concentration.*

*From Item 12 in reference list

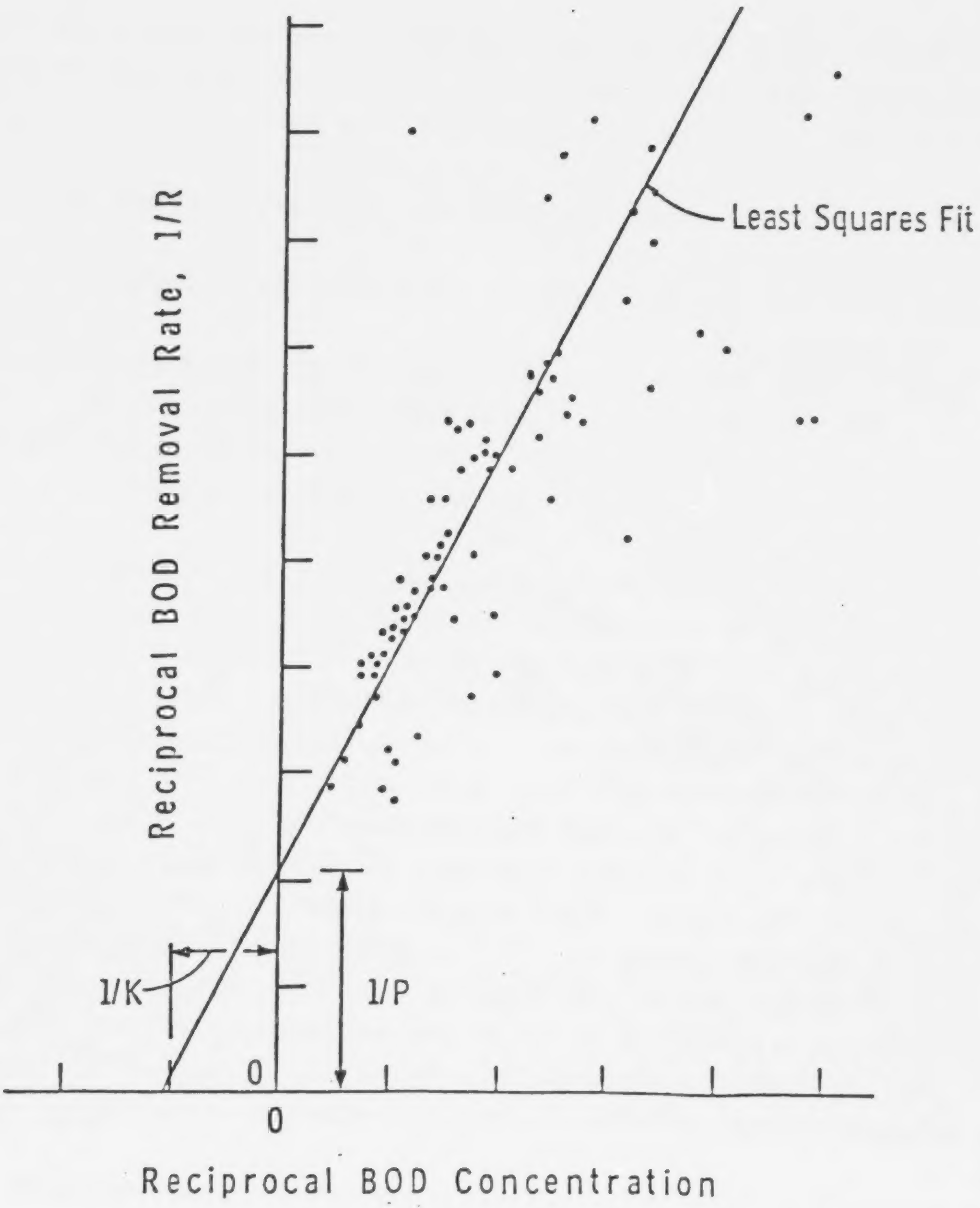

Figure 8-31(b). Reciprocal BOD removal rate versus reciprocal BOD concentration.*

*From Item 13 in reference list

Figure 8-31(c) is a plot of measured BOD versus BOD predicted by the equation. The fit indicates that this model, similar to the Monod bio-kinetic model, represents the data well.

Chevron believes (13) that RBC units for refinery service should be designed so that:

(a) The overall BOD removal rate is 0.002-0.003 lb/day/square foot.

(b) The BOD loading applied to the first RBC stage should not exceed 0.012-0.015 lb/day/square foot.

Chevron states that these factors, though tentative, should avoid excess and/or anaerobic bio-growth and mechanical overload.

Phillips Petroleum (15) found oxygen transfer to be essentially independent of disc rotational speed above 1 to 2 rpm (Figure 8-32) and found that supplemental in-tank aeration would only be necessary if soluble COD exceeded 200 mg/l. Phillips also established that hydraulic loading up to 1 gpd/square foot was low enough to meet permit limits of BOD = 39, COD = 190 and phenols = 0.28 mg/l. Whereas Chevron reported 0.002 to 0.003 lb BOD/day/square foot organic removal, Phillips found 0.0011 to 0.0042 lb COD/day/square foot organic removal. Phillips also found that whereas nearly all soluble BOD could be removed over a wide range of hydraulic and organic loadings, at low loadings nitrification resulted in up to 99 percent ammonia removal but at high loadings ammonia removal fell to zero. Figure 8-33 shows the ammonia removal efficiency versus hydraulic loading.

High hydraulic and organic loadings resulted in an increase in COD in the plant effluent even though soluble BOD removal remained high. Figure 8-34 shows soluble COD removal versus RBC stage for four pilot studies and one full-scale study. The effect of higher loading is clear although the full-scale plant (curve E) performance fell below the performance of the pilot

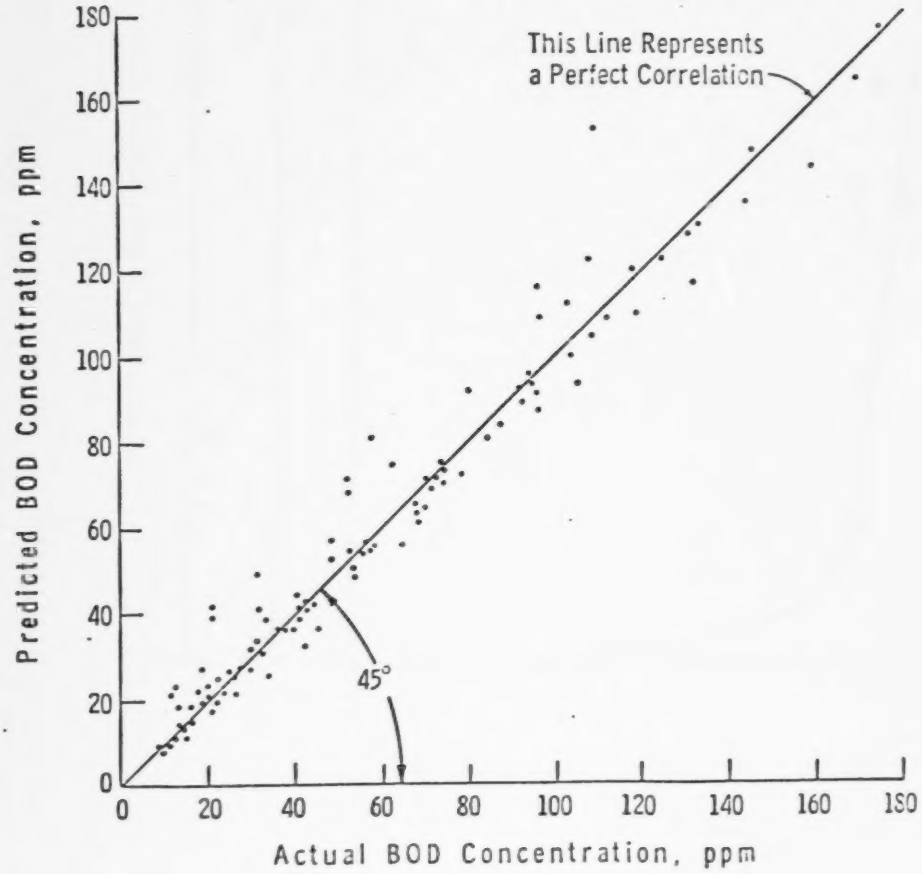

Figure 8-31(c). Accuracy of RBC model: predicted versus actual BOD*

*From Item 13 in reference list

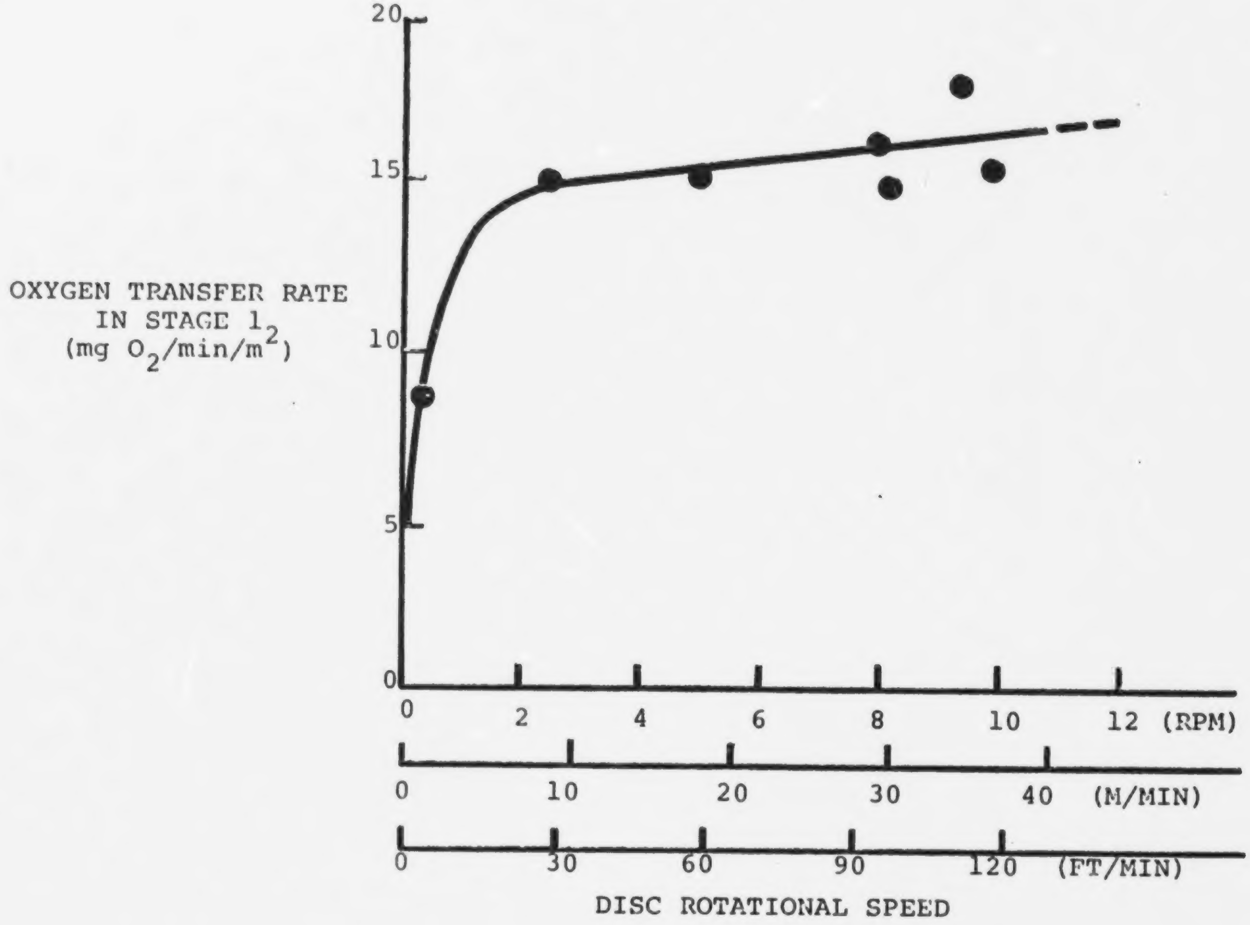

Figure 8-32. Relationship of oxygen transfer rate in the first stage of the RBC pilot unit versus rotational speed.*

*From Item 15 in reference list

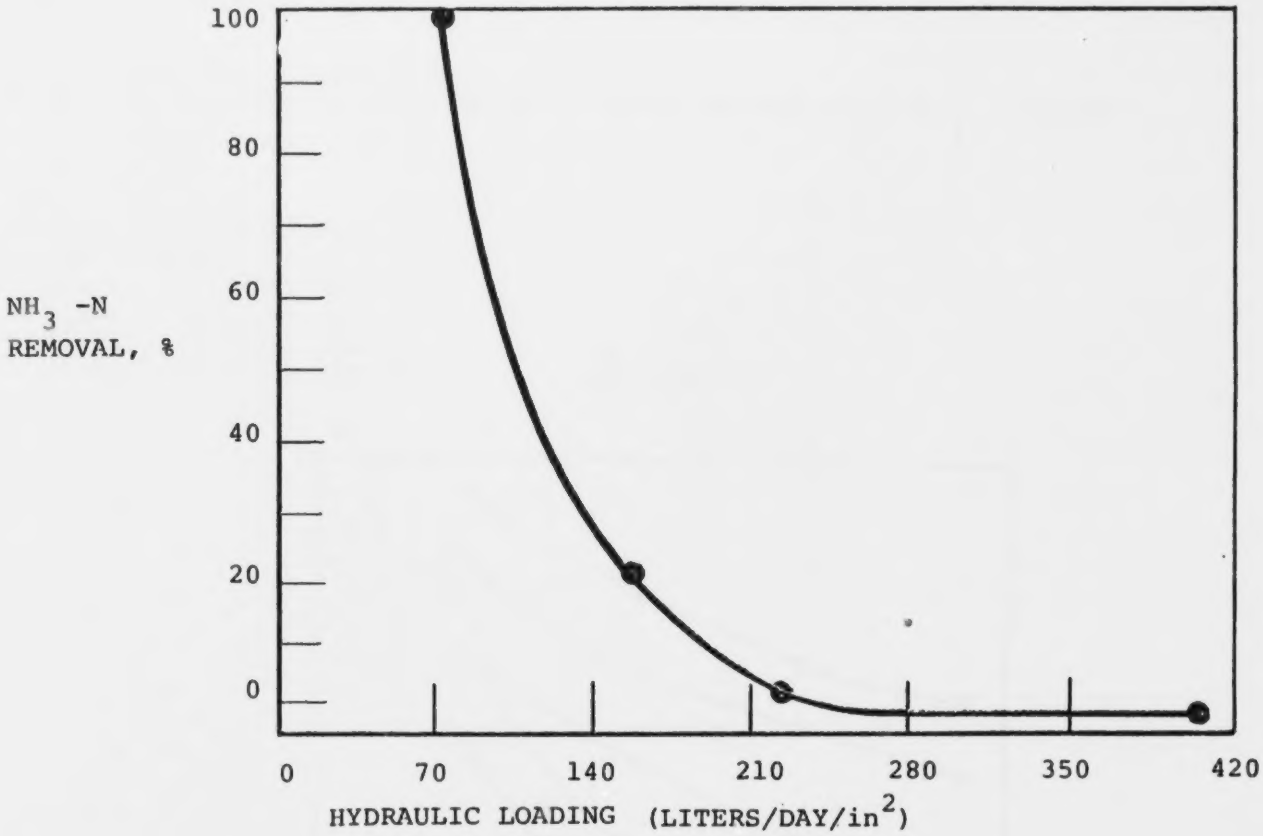

Figure 8-33. Ammonia nitrogen removal versus hydraulic loading in the RBC treatment unit.*

*From Item 15 in reference list

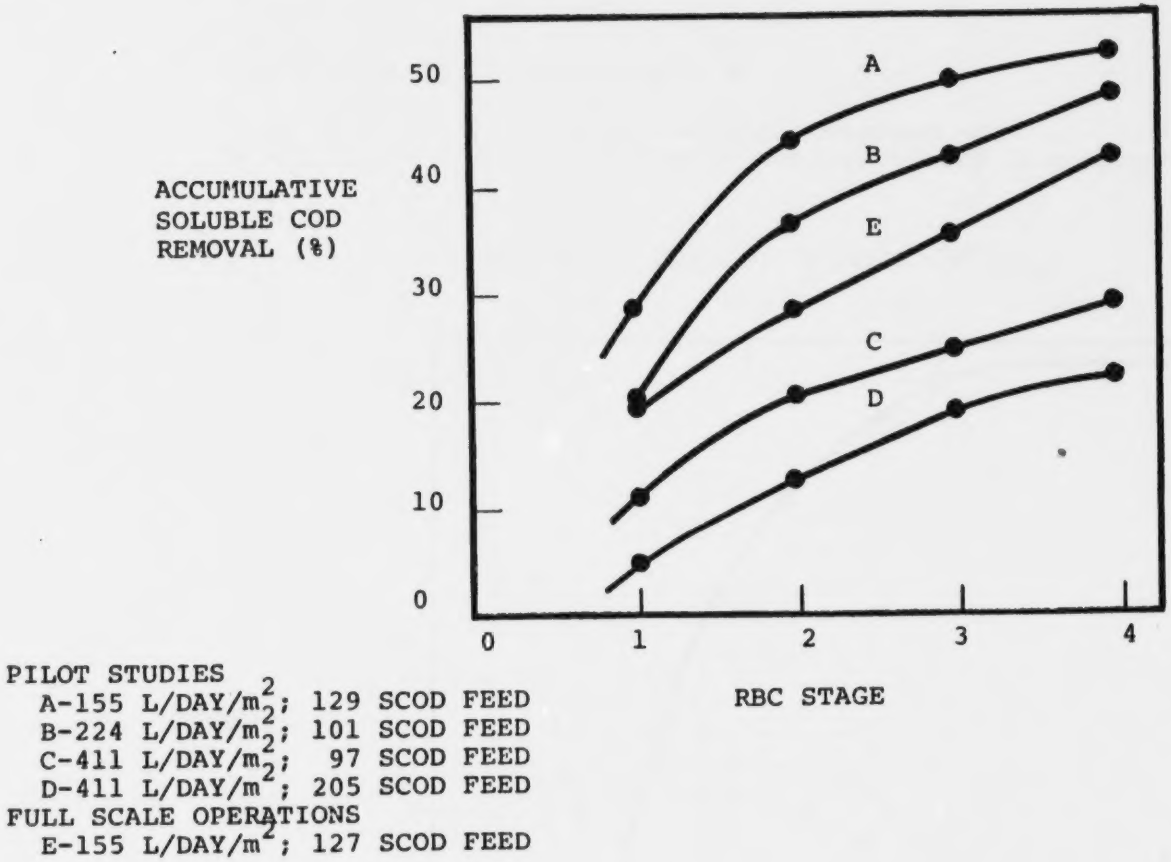

Figure 8-34. Comparison of soluble COD removal versus hydraulic loading through the RBC unit.*

*From Item 15 in reference list

plant (curve A) operating at the same loading. Figures 8-35 and 8-36 show, respectively, soluble BOD and COD removal efficiency versus influent BOD and COD concentration. Curves are shown for both pilot plant and full-scale operations.

Hormel(16) reports on four selected pilot RBC refinery studies, including the Chevron study in the case history. They also tabulate known refinery and pilot RBC studies. The tabulated studies had influent phenol from <1 to 500 mg/l and influent ammonia from 1 to 200 mg/l. Several were intended to "remove" ammonia.

The reported Hormel study for Vancouver, B.C. was designed for side-by-side comparison of pilot air activated sludge (AAS) and RBC units. Three categories of wastewaters were compared: oily, high-soluble organic and phenol-bearing. Both units satisfactorily reduced 47 to 375 mg/l oil influent to 15 to 31 mg/l in the effluent. AAS was more efficient than the RBC for soluble organics, and both removed over 99 percent of the 100 mg/l inlet phenol. In another study by Texaco in West Tulsa, Okla. on a dilute refinery waste (BOD <100), about 90 percent oil and phenol removals from average influent concentrations of 27 and 0.8 mg/l, respectively, were reported. Energy requirements for 1,000 gpm (1.44 MGD) RBC, AAS and aerated lagoon plants are summarized in TABLE 8-33. RBC at 71 hp is about one-half either of the others.

RBC Limitations--

RBC's are subject to several limitations involving mechanical considerations, shock loading, toxic substances and oil (13, 14, 16).

Care is required to ensure that the discs rotate and are immersed and that shaft bearings are lubricated. Various physical damages can arise from weight shifts when the bio-mass sloughs off the

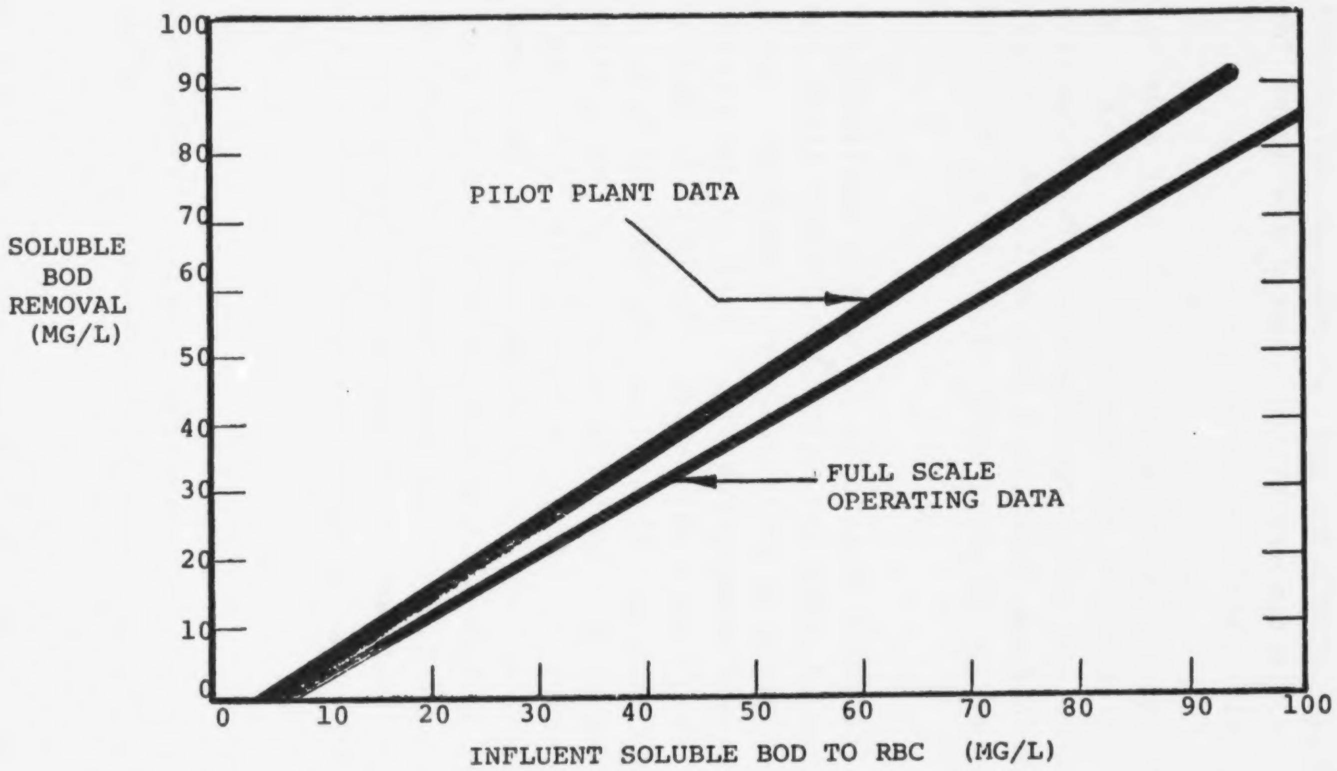

Figure 8-35. Soluble BOD removal efficiency for the RBC unit.*

*From Item 15 in reference list

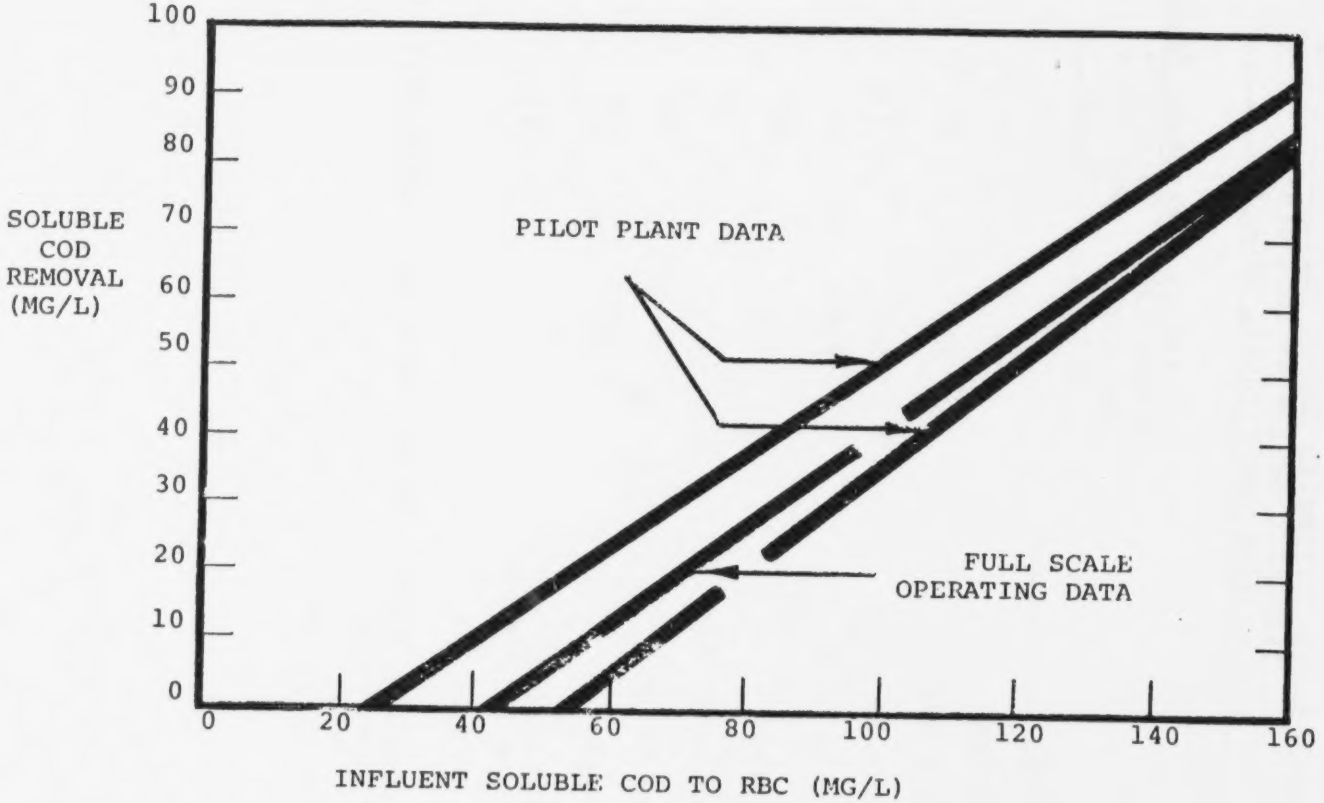

Figure 8-36. Soluble COD removal efficiency for the RBC unit.*
*From Item 15 in reference list

TABLE 8-33. ESTIMATED ENERGY REQUIREMENTS FOR INDICATED DESIGN AT 1000 GPM* (TEXACO, WEST TULSA, OKLA.)

	Continuous Horsepower Requirements		
Item	Recommended Rotating Disk	Alternate Activated Sludge	Existing Aerated Lagoon
Influent Pumps	5	5	5
Equalization Basin	30	30	15
Air Flotation Unit	--	28	--
Rotating Disk Unit	20*	--	--
Aeration Basin	--	40	135
Secondary Clarifiers	1	5	--
Multimedia Filters	--	20	--
Aerobic Digester	15	15	--
Total Estimated Continuous Horsepower	71	143	155

Note: Operating horsepower may be slightly less than indicated installed horsepower.

*From Item 15 in reference list

disc if rotation stops or the level in the RBC treatment tank falls too low.

Although no refinery had observed the effects of a shock BOD load (14) on RBC's, one chemical plant operator reported that a rapid sixfold BOD increase raised the effluent to 15 to 20 mg/l from its normal 2 to 5 mg/l. Recovery was reported in 24 hours.

With regard to toxic substances, one refiner reported that several weeks' caustic water accumulation introduced into the oily drain stripped the growth from the discs and 5 to 6 days were required for recovery. Another refiner found that 10 mg/l sulfide caused a white growth to develop on the first two stages of a four stage RBC and interfered with BOD removal and biomass settling. Reduction of sulfide to 1 mg/l allowed recovery in 1 to 2 weeks. A spill of thousands of pounds of sulfuric acid occurred (13) at Chevron's Refinery "B", causing pH to fall to 2 to 3 in the biological system. Although the RBC units sloughed off tons of bio-solids and temporarily overloaded the clarifier, the RBC's recovered to nearly full phenol activity in 24 hours.

Chevron reports (14) two experiences with oil. One refiner fed 1 percent oil to RBC's for two days due to a skimmer failure. Four to five days were required for recovery. The second refiner fed 100 to 200 mg/l oil for extended periods. Most of this oil passed through the RBC's and caused settling problems in the clarifier. In studies (16) that conflict with Chevron experience, Hormel reported oil reduction of 50 percent to over 90 percent in the range 12 to 375 mg/l influent oil.

RBC Costs--

Chevron (14) found that activated sludge could be cheaper than RBC's if wastewater was concentrated (above about 600 mg/l BOD), if the waste load exceeded about 15,000 lb/day BOD or if the activated sludge effluent requirement did not require filters to satisfy regulations. Hormel (16) found RBC energy needs to be about half those for other bio-processes on a comparable basis for a low-BOD 1.44 MGD design.

RBC Case History at Chevron Refinery "B"--

Refinery "B" is a 168,000 barrel per day plant that has fairly liberal effluent requirements to meet although they are more stringent in winter than the rest of the year. The RBC system at refinery "B" is the largest installed RBC unit treating refinery wastes. TABLE 8-34(a) shows the effluent requirements. TABLE 8-34(b) shows RBC performance for BOD and TABLE 8-34(c) shows the RBC clarifier performance for suspended solids (SS) for this installation. The flow is up to 3,000 gpm or 4.3 MGD (including 1,000 gpm storm flow)and after flotation is equalized in a tank of 12 hours' detention time. Pilot tests indicated 1.8 million square feet of RBC area was needed and 18 standard Autotrol 100,000 square foot "Bio Surf" units (manufactured by Autotrol Corporation) were employed in 5 parallel trains. Chevron experience indicated 3 to 4 units in series provide nearly optimum use of the area commensurate with the required removals. Each covered unit has a 7-1/2 hp drive and one train includes weigh cells on each shaft to determine the weight of bio-mass in each unit. TABLE 8-34(b) includes weigh cell measurements. These are quite valuable to detect toxic spill bio-mass destruction before it becomes catastrophic.

The initial RBC inlet BOD was 178 mg/l (period A) or 86 mg/l (period B). The combined RBC-clarifier outlet BOD was 34 mg/l and 18 mg/l for these same periods and thus combined removals were about 80 percent in each case.

Case History of Chevron Refinery "A" Pilot Tests--

Pilot tests were made at refinery "A", a 40,000 barrel per day refinery with existing API separator and pond system. TABLE 8-35

TABLE 8-34(a). EFFLUENT REQUIREMENTS AT REFINERY B, CHEVRON U.S.A.*

	30-Day Average (1)			
	December-March		April-November	
	ppm (2)	Lb/Day	ppm (2)	Lb/Day
BOD (5 day)	38	1,362	50	1,772
TOC	84	2,995	109	3,900
TSS	25	892	33	1,166
Phenol	0.19	6.9	0.27	9.7
NH_3 (As N)	16	578	23	805
Sulfide	0.16	5.6	0.22	7.8
Oil and Grease	11	403	15	525
pH	6-9		6-9	

(1) Requirements change according to seasonal asphalt production. An allowance for 1,000 gpm of storm water is included in these figures.

(2) ppm based on 2,970 gpm flow rate.

*From Item 13 in reference list

TABLE 8-34(b). INITIAL RBC PERFORMANCE AT REFINERY B, CHEVRON U.S.A.*

Period	Flow Rate gpm	BOD Reduction BOD (5-day), ppm RBC Inlet	 Clarifier Outlet
A	3,000-4,000	178 (117-241)	34 (23-44)
B	3,000-4,000	86 (65-100)	18 (14-25)
Design	3,000	165	38

Period	Weigh Cell Measurements, Lb Stage No. 1	 2	 3	 4
A	40-50,000	30,000	25,000	25,000
B	27,000	25,000	23,000	20,000

Values in parentheses show the range of data.

TABLE 8-34(c). INITIAL CLARIFIER PERFORMANCE AT REFINERY B, CHEVRON U.S.A*

Flow Rate, gpm	Suspended Solids, ppm Clarifier Inlet	 Clarifier Outlet
3000-4000	130 Average (20-360)	26 Average (9-40)
3000 Design	100 Design	20 Design

No. of data points - 52.
Values in parentheses show the range of data.

*From Item 13 in reference list

TABLE 8-35. PILOT BIOLOGICAL TESTS AT REFINERY A, CHEVRON U.S.A.*

RBC Units

A. Autotrol Biosurf Unit

15-In. Dia. Corrugated Polyethylene Media: 250 Sq. Ft. Surface Divided into Four Equal Stages

Clarifier: Hopper Bottom, 1.8 Sq. Ft. Overflow Area

B. Hormel RBC Unit

48-In. Dia. Flat Polystyrene Foam Disks: 800 Sq. Ft. Surface
Divided into Four Equal Stages

Clarifier: Hopper Bottom, 3.7 Sq. Ft. Overflow Area

Activated Sludge Unit

1,000-Gal. Clow Package Unit with Diffused Air Aeration

Clarifier: Hopper Bottom, 10 Sq. Ft. Overflow Area

Test Unit	Operating Conditions	BOD ppm(1) In	BOD ppm(1) Out	BOD Removal, %
Biosurf	2.0 Gal./Day/Sq.Ft. 85° F Inlet 75° F Outlet	103 (40)	60 (11)	42 (73)
RBC Unit	Same as Biosurf	103 (40)	47 (12)	54 (70)

(1) BOD values are averaged data for 5-day tests. Values in parentheses are soluble BOD's, others are for settled samples.
(2) Activated sludge unit never successfully started up.

*From Item 13 in reference list

describes the two pilot RBC units tested and compares conditions and pilot results for Autotrol Bio-Surf and the Hormel RBC units. Chevron concluded that the two RBC's were essentially equal in performance at 70 percent soluble BOD removal.

Case History at Chevron Refinery "C"--
Refinery "C" had ponds, but poor winter performance made upgrading necessary. Four 38,000 square foot Hormel RBC units were installed downstream of pond number 1. This RBC was the first installation in the U.S. for refinery waste treatment. Due to a decreasing waste load this system was underloaded and slime growth was thin. Side-by-side pilot and full scale tests were made to check for scale-up effects, but such effects were not found with this dilute feed operation. TABLE 8-36 summarizes the results of these tests and shows that approximately 30, 30 and 40 percent removals were obtained for BOD, TOC and phenols, respectively. Note that phenols were only about 130 parts per billion in the feed.

Activated Carbon Enhanced Activated Sludge (ACEAS)--
Activated carbon enhanced activated sludge processes involve a synergism between fluidized bed activated sludge and ordinary activated carbon wastewater treatment.

In ACEAS processes, aeration tanks receive primary treated wastewater, powdered activated carbon and secondary clarifier return sludge. In the aeration tank the bio-mass grows adherent to the individual carbon particles similar to the growth in other fluidized bed bio-reactors. However, the substrate is not inert, like sand, but adsorbent carbon. The adherent bio-mass exhibits the chemical and toxic shock resistance found in the other fixed film devices (trickling filters, fluid beds and rotating discs).

TABLE 8-36. PILOT VERSUS FULL-SCALE RBC UNITS AT REFINERY C, CHEVRON U.S.A.*

	BOD, ppm	TOC, ppm	Phenol, ppb
Full-Scale RBC			
Feed	47 (35-65)	41 (35-53)	130 (98-150)
Effluent	32 (36-44)	31 (21-44)	76 (51-95)
Reduction, %	32	24	42
Pilot RBC			
Feed	47 (35-60)	41 (35-48)	134 (115-159)
Effluent	33 (21-39)	29 (18-39)	80 (48-136)
Reduction, %	30	29	40

Average values are shown, with data range in parentheses.

Operating Conditions - Hydraulic Loading: 6.8 Gal./Day/Ft^2
Inlet Temperature: 55° F (Typical)
RBC Media:
Full-Size Unit: 4 x 38,000 Ft^2, 11-Ft Dia. Disks
Pilot Unit: 4 x 2,000 Ft^2, 4-Ft Dia. Disks
Peripheral Media Speed (Both Units): 51 Ft/Min.

*From Item 13 on reference list

In addition, the carbon may not only adsorb and desorb various substances as the waste concentration varies (18), but also serves to retain resistant organics in contact with the bio-mass for the full sludge age (18, 19), not just the hydraulic retention time. Finally, in the secondary clarifier the carbon-weighted bio-mass settles readily where it is split into return and waste activated sludge as in ordinary activated sludge systems.

Waste sludge from ACEAS processes may be sent to disposal or it may be processed to recover activated carbon. DuPont stockpiled literally hundreds of tons of carbon containing sludge from the PACT process at their Chambers works while awaiting repairs to the reactivation furnace (summer/fall 1977) and experienced no difficulties other than those caused by rain and ice. Sludge settling, handling and dewatering may be enhanced by carbon (18).

Both conventional dry thermal (19) and wet oxidation (20) carbon regeneration methods have been used for carbon recovery. Regeneration losses are 5 to 15 percent per pass. Regeneration is an economic necessity for systems with conventional sludge ages of 5 to 15 days and where the carbon makeup rates needed are 50 to 450 mg/l to obtain aeration tank carbon levels of 500-5,000 mg/l. Based on refinery studies (21, 22), carbon recovery is not an economic necessity for systems with less than about 50 mg/l makeup rates based on influent flow rate.

The relationship between aeration tank equilibrium carbon concentration C, carbon dose rate C_i, hydraulic retention time and sludge age is:

$$C = C_i \text{ (SA/HRT)}$$

where SA = sludge age
HRT = hydraulic retention time

Thus aeration tank carbon levels of 500-5,000 mg/l, comparable to the conventional SA high dose rate (50-450 mg/l) units described above, can also be obtained at much lower dose rates if SA's are extended to 20 to 50 days.

The following table shows the variation in equilibrium carbon concentration for several carbon dosage rates and sludge ages. The hydraulic retention time is one day in all cases. The assumption is made that carbon losses in the effluent are accounted for and are lower than the dose rate.

Equilibrium Carbon Concentration (C) at HRT = 1 Day

Dose Rate	SA, days		
C_i, mg/l	10	25	50
10	-	-	500
50	500	1,250	2,500
200	2,000	5,000	-

Carbon doses as low as 10 to 15 mg/l have been used in some pilot tests and full scale demonstrations (22, 23). Analysis may show that systems designed for long SA's and with dosage rates below about 50 mg/l are economically attractive without carbon recovery: the increased capital required for the long SA may be more than compensated for by the elimination of costs of carbon recovery.

ACEAS Applications--

Some of the physical properties of activated carbon are important in the application of ACEAS processes. The most notable are surface area (21, 22) and density (18, 23). Reference 22 is an important report by the API discussing the ACEAS process application to refinery and petrochemical waste treatment. Reference 22 and reference 23, an oil company paper, both discuss the advantages of experimental high surface area activated carbons (Amoco PX Series). The API study found that the high area carbon could be combined with a high SA to maintain 2,500 mg/l carbon in the aeration tank and COD removals were obtained comparable to those from add-on granular carbon columns. The study found incremental costs for COD removal below that obtained with filtered activated sludge were $0.61/lb COD for powdered carbon versus $3.19/lb COD for granular carbon (1977 costs for a 1 MGD typical refinery activated sludge unit).

In some contrast to these studies, but with similar results, Adams and DeJohn (24) found carbon density to be important to ACEAS processes because high-density carbon so reduced clarifier (effluent) losses that high SA's were feasible. Reference 24 has a good discussion on the benefits and mechanisms of powdered carbon in activated sludge units and presents four refinery case histories of ACEAS operations.

ACEAS processes have shown great promise with refinery wastewaters, are especially applicable to toxicity and high concentration problems, and should be tested with coal conversion wastewaters in pilot and demonstration units.

References for Biological Oxidation--

1. EPA 625/1-71-004a, "Upgrading Existing Wastewater Treatment

Plants," October 1974. P. 4-4, 4-23, 5-3 and 5-20. 870*

2. EPA 600/7-77-065 "Water Conservation and Pollution Control in Coal Conversion Processes," Water Purification Associates, June 1977. P. 401, 399, 396, 393, 369 and 366. 480*

3. Adams, C.E., Jr., Stein, R.M. and Eckenfelder, W. W., Jr., "Treatment of Two Coke Plant Wastewaters to Meet Guideline Criteria," Proceedings 29th Purdue Industrial waste Conference, May, 1974. P. 864-880.

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5. Gould, M.S., Roy, A.R., and Genetelli, E.J., "Pretreatment of a Highly Organic Industrial Wastewater - Case History," Proceedings of the 29th Purdue Industrial Waste Conference, May 1974. P. 889-896.

6. Cooke, R., and Graham, P.W., "Biological Purification of Effluent from a Lurgi Plant Gasifying Bituminous Coals." International Journal of Air and Water Pollution, Pergamon Press 1965, Vol. 9. P. 97-112. 697*

7. PERC/RI-77/13 "Treatability Studies of Condensate Water from Synthane Coal Gasification." November 1977. 797*

8. Davis, G.M., Reap, E.J., and Koon, J.H., "Treatment Investigations and Process Design for the H-Coal Liquefaction Wastewater." Unpublished report for Ashland Oil by Aware, Inc. December 1976. 678*

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9. Adams, C.E., "Treatment of a High Strength Phenolic and Ammonia Waste Stream by Single and Multi-Stage Activated Sludge Processes." Proceedings of the 29th Purdue Industrial Waste Conference, May 1974.

10. Kostenbader, P.D., and Flecksteiner, J.W., "Biological Oxidation of Coke Plant Weak Ammonia Liquor," JWPCF (41) 2 Part 1, February 1969.

11. EPA-R2-73-167, "Biological Removal of Carbon and Nitrogen Compounds From Coke Plant Wastes." April 1973. 800*

12. Parsons, W.A., and Nolde, W., "Applicability of Coke Plant Water Treatment Technology to Coal Gasification." presented at EPA Symposium, Hollywood, Florida, September 1977. 958*

References for Rotating Biological Contactors--

13. Davies, B.T., and Vose, R.W., "Custom Designs Cut Effluent Treating Costs. Case Histories at Chevron U.S.A., Inc." Presented at 32nd Purdue Industrial Waste Conference, May 1977. 653*

14. Knowlton, H.E., "Why Not Use a Rotating Disc?" Hydrocarbon Processing, September 1977, p. 227-230.

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16. Flann, G.E., and Gerhard, R.E., "Use of The Rotating Biological Surface For Refinery Wastewater Treatment." Presented at 69th Annual Meeting AIChE, Chicago, November-December 1976. *545

17. McAliley, J.E., "Pilot Plant Study of a Rotating Biological Surface for Secondary Treatment of Unbleached Kraft Mill Waste." Tappi, Vol. 57, No. 9 (1974). P. 106. 780*

References for Activated Carbon Enhanced Activated Sludge--

18. Flynn, B.P., Robertaccio, F.L., and Barry, L.T., "Truth or Consequences: Biological Fouling and Other Considerations in The Powdered Activated Carbon-Activated Sludge System." Proceedings of the 31st Purdue Industrial Waste Conference, May 1976. 931*

19. Adams, A.D., "Powdered Carbon: Is It Really That Good?" Water and Wastes Engineering, March 1974. 928*

20. Flynn, B.P., and Barry, L.T., "Finding a Home For The Carbon: Aerator (Powdered) or Column (Granular)." Proceedings of 31st Purdue Industrial Waste Conference, May 1976. 931*

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Filtration of Effluent from Biological Treatment

Although filtration is a fundamental unit operation with wide applications, this description is concerned only with liquid-solid filtration as applied to the polishing of coal conversion wastewater secondary (biological) effluent. The main advantage of filtration in this use is that it provides positive suspended solids (SS) control following the secondary clarifier. The EPA Process Design Manual (1) on upgrading existing municipal wastewater treatment plants has a good discussion of effluent polishing techniques, including filtration, in Chapter 7. The manual describes three filtration applications for polishing secondary effluents:

- Directly after the clarifier. Biological flocs are usually strong and multimedia or granular media filters that provide depth filtration perform well.
- After chemical clarification. The flocs are frequently weak and require sand or uniform-media surface filters for good separation.
- Following in-line chemical injection. As with chemical clarification, the flocs are usually weak and require surface filters.

Mechanical considerations necessary for all types of biological filtering include thorough backwashing facilities with air scour or hydraulic jets plus normal upflow wastewater and periodic shock chlorination. Backwash flow equalization may be necessary in small plants. Backwash should be returned to the feed equalization tank or directly to the aeration tanks of the biological oxidation system. If filter effluent is used for backwash, backwash storage tanks must be provided.

Because of the developing state of coal conversion plants and their wastewaters, pilot plant filtration studies will be necessary for economic design. A major factor influencing efficiency and economics will be the consistency of biological treatment expected for these wastewaters (poor stability has been observed in coke liquor biological treatment). Piloting will provide the needed information on effects of variability as well as accurate design data for the effects of influent solids concentration on run length for various loadings.

Reference 2 gives a good description of pilot filtration studies made at the Metropolitan Denver AS plant. Filterability studies on filter effluent turbidity versus bed depth profiles are provided for three run lengths for each of the three filtration applications. Effects of bed-depth, media type and media size were investigated as well. The media used were coal, sand and garnet. A second set of studies provided profiles for turbidity versus bed depth with coal and sand in dual- and mixed-media modes. They concluded that dual coal/sand media filtered AS effluents directly to 1 to 2 mg/l SS with runs over 20 hours long at 6 gpm per square foot. Alum coagulation definitely decreased filterability, requiring finer filter media for removal.

In their paper "Applying Coke Plant Water Treatment Technology to Coal Gasification," Parsons and Nolde (3) say that although the present trend at coke plants is to AS treatment, limitations include inconsistent discharges of SS, thiocyanates,and cyanides. From coke plant AS experience they expect coal gasification sour water AS plants to have 60 to 200 mg/l SS in the settled effluent (about 5 times higher than municipal AS effluents). These high levels of SS will require that filters be relatively lightly loaded to achieve reasonable run times, and a relatively large flow of backwash will be produced. The following table shows some examples of direct filtered AS effluents for municipal wastes (1).

SELECTED MUNICIPAL ACTIVATED SLUDGE EFFLUENT DIRECT FILTRATION EXPERIENCE*

Plant Location	Load Rate gpm/ft^2	Influent SS mg/l**	Effluent SS mg/l**	Run Length, hrs
Louisville, KY.	3.4	27 (11-32)	3 (1-4)	
Ann Arbor, Mich.	6	42 (28-126)	5 (1-17)	
Philomath, OR.	5	165 (30-2,180)	5 (1-20)	(15-24)
Hanover Park, IL.	2.2	14	7	

* From Item 1, pp. 7-20, in reference list

** Range shown in parentheses

References--

1. EPA 625/1-71-004a, "Upgrading Existing Wastewater Treatment Plants," October 1974. P. 7-12 to 7-23, 7-36 to 7-37. 870*

2. Maxwell, M. J., Linstedt, K. D., Work, S. W. and E. R. Bennett, "Making Optimum Use of Filter Media in Wastewater Filtration," Water and Sewage Works. Dec. 1977.

3. Parsons, W. A. and W. Nolde, "Applicability of Coke Plant Water Treatment Technology to Coal Gasification," presented at EPA Symposium, Hollywood, Florida, Sept. 1977. 958*

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*Pullman Kellogg Reference File number

Sludge Handling and Disposal

Liquefaction processes and gasification processes that produce phenols, oils,and tars generate strong sour waters with high BOD and COD. These sour waters contain toxic and organic substances that may be economically treated for reuse or discharge in biological oxidation processes.

Generally, the sludge quantities produced by biological oxidation are directly proportional to the BOD or COD removed, are higher for more degradable wastes, decrease with increasing sludge age, and are less voluminous for fixed film processes than for suspended growth processes.

Coal conversion sour waters will require treatment both before and after biological oxidation. Pretreatment steps anticipated are degasification, API separation, inert dissolved gas flotation, phenol recovery, stripping,and equalization. Primary and secondary oil removal by API and flotation, respectively, are included in pretreatment because of the adverse effects of mineral oils on biological oxidation and other downstream processes. Sludge production is affected by oils because the oils coat and adsorb onto the sludge flocs and reduce their settleability. Posttreatment steps include chemical coagulation and filtration plus various other processes like reverse osmosis, depending on the end use requirements.

All biological oxidation processes produce biomass growth in a reactor system. The excess growth leaves the reactor with the effluent and is usually settled out by gravity in a secondary clarifier. Clarifier underflow or solids flow is usually quite dilute, perhaps 0.5 to 1.5 percent solids for activated sludge systems. In the activated sludge system the clarifier underflow is split into the return sludge stream and the waste activated sludge stream (WAS). The WAS is the sludge production from biological oxidation treatment and must be further processed for separation and disposal. Sludge handling is responsible for 30 to 40 percent of the capital cost of a municipal treatment plant and about 50 percent of the operating cost.

Wastewaters from coal conversion processes that produce phenols, oils and tars (p/o/t) may be treated by biological oxidation. P/o/t-producing conversion processes include all low-severity operations such as Lurgi, Synthane and Hygas gasification as well as SRC, H-Coal and Synthoil solvent-based liquefaction processes. After proper pretreatment, wastewaters from these processes may be processed by activated sludge biological treatment.

The amount of waste sludge produced from activated sludge systems may be calculated by (1,2):

$$\text{Net Sludge, lbs} = a(L_R) - b(S_A)$$

where a = constant for yield
b = constant for endogenous respiration
L_R = pounds BOD removed by the process
S_A = mass of mixed liquor volatile suspended solids (MLVSS)

Constants a and b may range from 0.3 to 0.7 and 0.03 to 0.14, respectively, and thus for many systems the net sludge production is <0.5 pounds per pound of BOD removed.

For gasification processes producing p/o/t, published estimates (3) of biosludge production ranged from a low of 29,000 lb/day for Lurgi with phenol recovery to a high of 192,000 lb/day for IGT Steam/Iron HyGas without phenol recovery for commercial size (250 billion Btu per day of product gas) designs. Unrecovered phenols were destroyed by biological oxidation in the latter case. Unpublished estimates by Kellogg and vendors for Lurgi gasification ranged from 17,000 to 24,000 lb/day with prior phenol recovery.

For liquefaction processes, a published paper (4) on pilot wastewater treatment studies for H-Coal gave sludge coefficients of a = 0.48 and b = 0.03 on a volatile solids basis for conventional activated sludge. Based on these coefficients, a commercial plant with similar waters and about 40,000 lb/day BOD removal would produce 17,000 lb/day waste activated sludge.

Handling and Disposal--

Waste activated sludge and other biological sludges are unstable and must be processed further to prepare them for storage and disposal.

Biological sludges range from 93 to 99 percent water (5) as produced and waste activated sludge is 98.5 to 99.5 percent water. The dry solids content of these sludges typically contain about 2/3 volatile organics and will putresce if allowed to stand. The water must be reduced and the organic content must be stabilized before these sludges can be stored or land filled. Figure 8-37 shows the general process sequence in municipal sludge handling: thickening, stabilization, conditioning, dewatering, heat drying, reduction and final disposal. Steps with the dashed lines through the middle may be bypassed for some sequences.

The most common stabilization practice for refinery sludges is aerobic digestion (6), usually preceded by thickening to reduce the hydraulic load. Thickener supernatant is returned to the activated sludge reactor or aeration tank. After digestion the stabilized sludge is dewatered, prior to final disposal, by filters, centrifuges or presses. Organic polyelectrolyte conditioners are usually added for most dewatering processes. The EPA "Process Design Manual for Sludge Treatment and Disposal," (5) is an excellent source of information and design data.

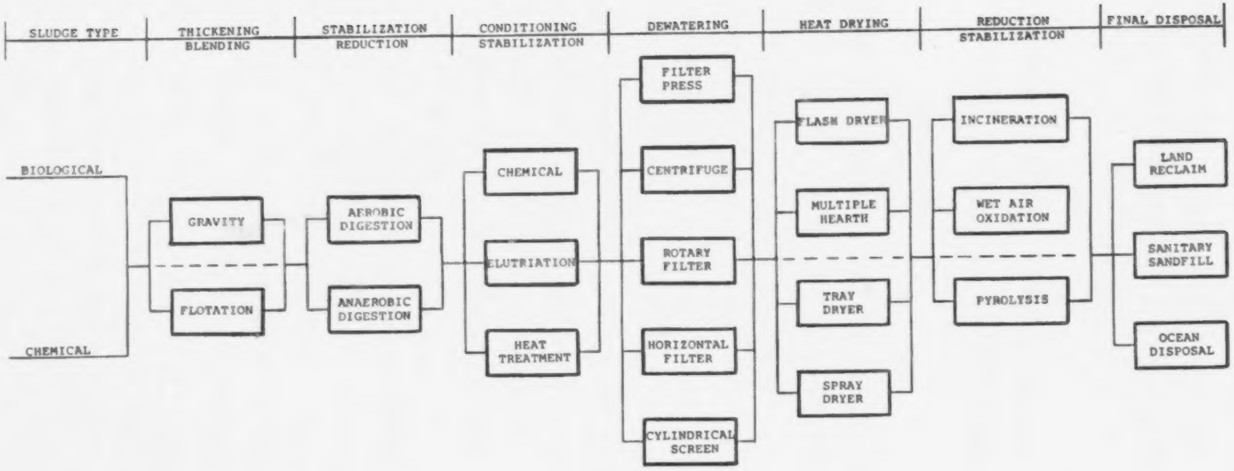
SLUDGE TYPE
THICKENING BLENDING
STABILIZATION REDUCTION
CONDITIONING STABILIZATION
DEWATERING
HEAT DRYING
REDUCTION STABILIZATION
FINAL DISPOSAL
BIOLOGICAL
CHEMICAL
GRAVITY
FLOTATION
AEROBIC DIGESTION
ANAEROBIC DIGESTION
CHEMICAL
ELUTRIATION
HEAT TREATMENT
FILTER PRESS
CENTRIFUGE
ROTARY FILTER
HORIZONTAL FILTER
CYLINDRICAL SCREEN
FLASH DRYER
MULTIPLE HEARTH
TRAY DRYER
SPRAY DRYER
INCINERATION
WET AIR OXIDATION
PYROLYSIS
LAND RECLAIM
SANITARY SANDFILL
OCEAN DISPOSAL

Figure 8-37.- Unit processes in sludge processing and disposal.

The most important common sludge characterization parameter is the Sludge Volume Index (SVI), defined as 1,000 times the volume occupied by the sludge layer in a one-liter sample, after 30 minutes settling in a liter graduate, divided by the initial solids concentration:

SVI = (Volume of sludge x 1,000)/(Initial suspended solids, mg/l)

For activated sludge plants a sludge with a SVI less than 100 is considered good settling whereas a sludge whose SVI is greater than 100 is often troublesome. Limited data from coal conversion wastewater pilot studies show that biosludges from coal conversion may be similar to biosludges from municipal and other industrial processes. Synthane pilot activated sludge units produced sludges with SVI's of 20 to 55 for a F/M range of ≤ 0.1 to 0.9 (7). H-Coal liquefaction pilot sludge had SVI's in the range of 70 to 90 for F/M from 0.03 to 0.23 (8). Both these sludges handled well.

Biological sludge can be thickened either by gravity or by air flotation (6) to 2 to 4 percent solids. Aerobic digestion will reduce volatile organic solids by 50 to 60 percent in 10 to 15 days' hydraulic detention. Conditioning with ferric chloride at 200-400 lb/ton of dry sludge solids or more will produce cakes of 10 to 16 percent solids from vacuum filtration, and up to 50 percent solids from pressure filtration. Both cakes are suitable for landfill. The pressure filtration cake should provide a net heat on combustion: no fuel need be added.

References--

1. Wilkinson, J.B., "Predicting Sludge Production From Refinery Activated Sludge Oxygen Uptake."

Purdue Industrial Waste Conference May 1976, p. 605. 931*

2. Eckenfelder, W.W., Jr., "Industrial Water Pollution Control," McGraw Hill, 1966, p. 162. 932*

3. FE 2240-5 "Factored Estimates For Western Coal Commercial Concepts," Interim Report, October 1976, to DOE and AGA by C.F. Braun & Co., Alhambra, Ca. 294*

4. Reap, E.J., Davis, G.M.; Duffy, M.J., and Koon, J.H., "Wastewater Characteristics and Treatment Technology For The Liquefaction of Coal Using H-Coal Process," Proceedings of the 32nd Purdue Industrial Waste Conference, May 1977. 654*

5. EPA 625/1-74-006. "Process Design Manual for Sludge Treatment and Disposal." USEPA Technology Transfer, October 1974. 868*

6. Adams, S, C.E., "Sludge Handling Methodology for Refinery Sludges," Proceedings of EPA Open Forum on Management of Petroleum Refinery Wastewaters, Editor F.S. Manning, University of Tulsa, Tulsa, Oklahoma, 1976. 828*

7. PERC/RI-77/13 "Treatability Studies of Condensate Water From Synthane Coal Gasification." November 1977. 797*

8. Davis, G.M., Reap, E.J., and Koon, J.H., "Treatment Investigations and Process Design for the H-Coal Liquefaction Wastewater." Unpublished report for Ashland Oil by AWARE, Inc., December 1976. 878*

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*Pullman Kellogg Reference File number

Carbon Adsorption

Activated carbon (AC) is a powerful adsorbent available in granules or powders. AC is commonly used in wastewater treatment as a granular material in packed beds. When powdered activated carbon (PAC) is used it is in suspended beds and treatment is followed by chemical or biological flocculation and settling. Granular activated carbon (GAC) is roughly twice as costly as PAC on a weight basis and thus GAC is commonly regenerated and reused whereas PAC is sometimes used once and discarded.

There are three broad-use categories of AC in wastewater treatment: physical-chemical (PC), activated sludge enhancement and tertiary treatment or polishing. AC, usually in granular form, can be used in conjunction with such other physical and chemical means as chemical clarification and multi-media filtration to provide relatively thorough wastewater cleanup via PC treatment (1). For activated sludge enhancement, PAC may be added to the aeration basins, as in DuPont's proprietary "PACT" system which is described separately. In tertiary treatment, GAC may be used after biological oxidation (biox) to reduce residual COD and trace organics.

The most common wastewater applications of AC adsorption use granular material in a series of fixed beds. Often only two beds in series constitute a "train" and therefore for reliability a minimum of two trains must be provided. In use, one bed in a train is on standby while one (or the rest) is adsorbing. At break-through, flow is switched to the standby and the spent bed is slurried or otherwise transported to regeneration. The empty unit is refilled with regenerated or new GAC and placed on standby. Fixed beds, either upflow or downflow, have both an advantage and a disadvantage in that fixed beds act as filters. An advantage, if suspended solids (SS) are low, is that fixed beds will retain SS and thus act as polishing filters as well as adsorbers. A disadvantage, if SS are high, is that physical plugging can precede carbon exhaustion and thus require excess carbon capacity and high backwash flows. Fixed beds also grow bio-slimes in many applications and either become plugged or generate H_2S or both, and both problems affect design and operation.

GAC can also be used in upflow expanded bed operations. The EPA manual on carbon adsorption (1,3-21*) recommends upflow expanded bed operation for H_2S problem control if the applied BOD exceeds 5 mg/l, which it surely will for coal wastewaters. The EPA manual is recommended for more design information and many municipal applications.

Activated carbons and waste adsorbability are characterized by batch isotherm and continuous column studies. Isotherm studies are used to find constants K and n in the Freundlich equation by log-log plots of X/M versus C_e. The Freundlich equation is:

$$X/M = K(C_e) \exp 1/n$$

Where X/M = Weight adsorbate adsorbed per weight carbon
C_e = Equilibrium adsorbate remaining, TOC or COD
K, 1/n = Empirical characterization constants.

Figure 8-38 is a plot of the Freundlich equation with slope 1/n and y-intercept K. Although Freundlich constants describe the adsorbability of pure compounds they are not very useful for complex wastes. They are useful for comparing carbons and have limited use for judging AC treatment effectiveness (3).

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*Item 1 in reference list, page 3-21.

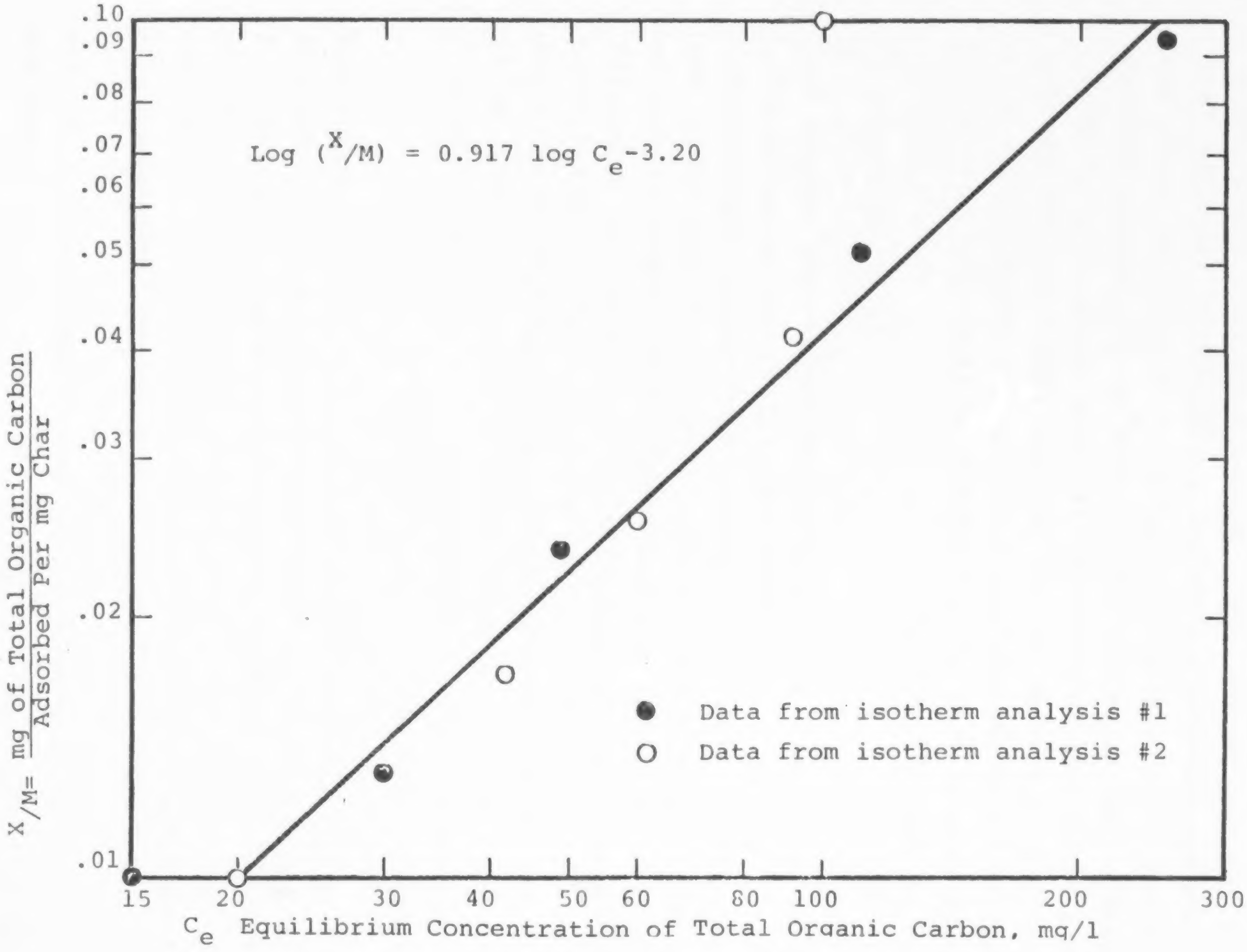

Figure 8-38. Montana char isotherm: Synthane biox effluent

(From Item 2 in reference list)

Continuous column studies are used to determine "breakthrough" curves by plots of effluent concentration versus wastewater throughput (in bed volumes, BV). Figure 8-39 is a breakthrough plot for a bio-treated petrochemical waste with a variable feed concentration (COD) showing the results of the column study and of a second column in series. The figure is interesting because it shows a rapid loss in adsorbability from column 1 to column 2. For example, at a flow of 50 bed volumes (BV), COD was reduced approximately 60 percent by the first column (from 650 to 260) but only 20 percent more by the second to (210) for an overall reduction of about 70 percent.

Limitations--

Adsorption decreases with decreasing molecular weight and aromaticity and with increasing solubility and polarity (4). TABLE 8-37 shows a range of 7 to 97 percent removal (adsorption) of a variety of pure compounds with a dose of 5 gm/l PAC. The tabulated data show that residual aromatics and phenolics will be much more efficiently reduced than will residual low molecular weight acids and alcohols. The effects of pH on adsorption depend on the adsorbate polarity: non-polar substances will not be affected by pH, adsorption of phenols and other organic acids will be enhanced by low pH and adsorption of organic bases will be enhanced by high pH.

Tests made on bio-treated petrochemical wastes by Lawson and Fisher (4) showed that the first bed in a two-bed train removed most of the readily adsorbable COD and the second column removed little additional COD. See Figure 8-39. Thus, some COD constituents resist adsorption and may constitute a refractory residual too large for release or use as boiler feedwater.

Suspended solids may foul packed carbon beds rapidly if over about 50 mg/l including bacterial (5,6). Free oil and grease foul

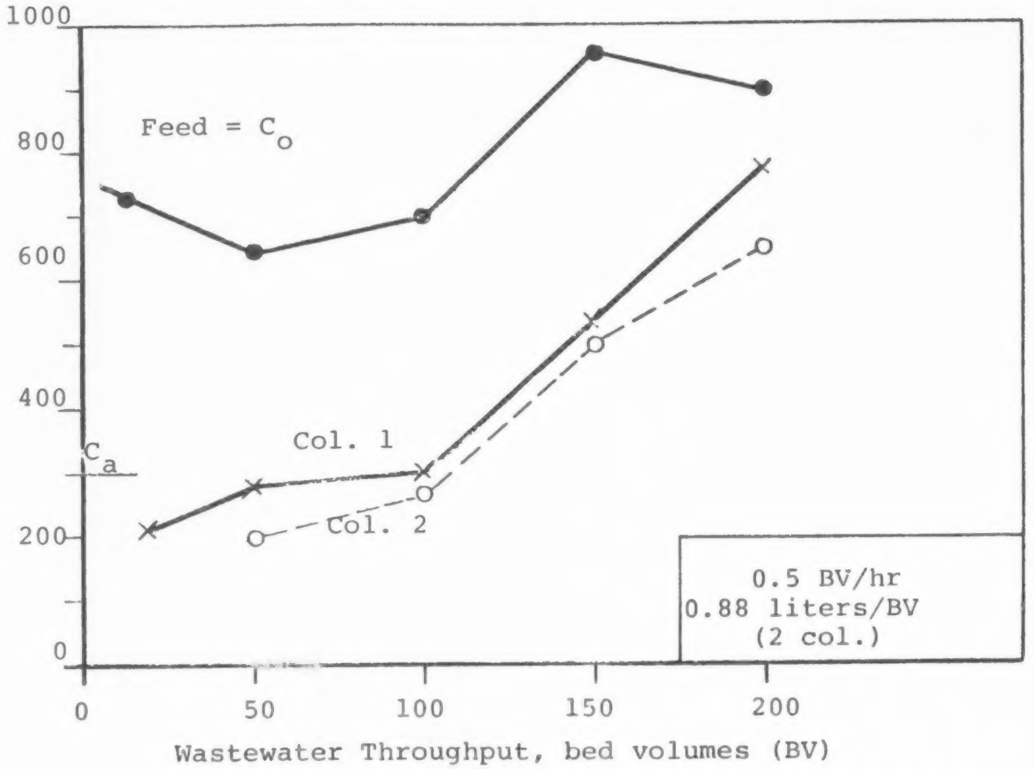

Figure 8-39. Breakthrough curve for Plant B bio-treated wastewater.*

*From Item 6 in reference list

TABLE 8-37. RELATIVE AMENABILITY TO ADSORPTION OF TYPICAL PETROCHEMICAL WASTEWATER CONSTITUENTS (a)

Compound at 1,000 mg/l Initial Concentration (b)	Percentage Removal of Compound at 5 gm/l Powdered Carbon Dosage
Ethanol	10
Isopropanol	13
Acetaldehyde	12
Butyraldehyde	53
Di-N-propylamine	80
Monoethanolamine	7
Pyridine	47
2-Methyl 5-ethyl pyridine	89
Benzene	95
Phenol	81
Nitrobenzene	96
Ethyl acetate	50
Vinyl acetate	64
Ethyl acrylate	78
Ethylene glycol	7
Propylene glycol	12
Propylene oxide	26
Acetone	22
Methyl ethyl ketone	47
Methyl isobutyl ketone	85
Acetic Acid	24
Propionic acid	33
Benzoic acid	91

(a) Item 4 in reference list.
(b) Benzene test at near saturation level, 420 mg/l.

carbon at 10 to 30 mg/l (5,6) and must be reduced by pretreatment. Ford (3) states that sour waters should not be treated by AC without proper pretreatment because of their tendencies to foul the beds by slime growth and to release H_2S. The EPA "Process Design Manual for Carbon Adsorption" (1) states that for applied BOD values substantially above 5 mg/l, upflow expanded aerobic beds should be used for H_2S control.

Applications--

Processes like Synthane and Lurgi that produce high phenols, oils, and tars (p/o/t) will require complete treatment with extraction, stripping and coagulation preceding biox treatment. GAC will be a useful tertiary treatment step to reduce residual organics but additional treatment will be required before the carbon beds, including filtration and ozonation. Ford (3) says ozonation may be a particularly effective pretreatment for adsorption. The Synthane PDU gasifier wastewaters have been subjected to a series of treatment processes culminating in adsorption with gasifier char (2). The char had a surface area of 325 square meters per gram, about 1/3 to 1/2 that of commercial carbons (1) used for wastewater treatment. With influent TOC of 80 to 490 mg/l, char TOC removals were 88 to 97 percent. TOC residuals were 35 mg/l or less.

Gasification processes like the Koppers-Totzek produce few organics and have low COD (7). They do have objectional quantities of some inorganics including cyanides, ammonia, hardness, SS and trace metals. Treatment might include chemical softening, filtration and ozonation or chlorination followed by GAC.

Liquefaction processes produce essentially CO_2-free sour water from dissolver off-gas condensates. These waters have ammonia associated with sulfides and phenols and are low in hardness and carbonate alkalinity. TABLES 8-38 and 8-39 show analyses for unstripped H-Coal and SRC sour waters. These strong waters will be subjected to a series of treatment steps including biological oxidation. TABLE 8-40 shows the effects of biological oxidation on an H-Coal PDU waste, yielding a residual soluble COD of 360 mg/l. Filtration or coagulation-clarification could be used to reduce SS from 60 mg/l before AC treatment but soluble COD would not be reduced by SS removal alone. Thus, ozonation and AC would be the next treatment processes after biological oxidation and filtration or clarification.

Case Histories--

A recent report (2) concerns treatability studies made on Synthane PDU waters and describes char adsorption of waters after biological oxidation treatment. The Synthane char had a surface area of 325 square meters per gram, about 1/3 to 1/2 that of commercial AC used in waste treatment. The waters studied were produced from a decant tank following the gasifier condensers. All waters were produced from a subbituminous "C" coal, Montana Rosebud. Pretreatment for biological oxidation included air-stripping ammonia at pH 11 to reduce ammonia to less than 500 mg/l and removal of tars, oils and grease (TOG) with 100 to 150 mg/l alum at pH 1.5-2.5. Effluent from pretreatment contained about 600 mg/l soluble TOG and was then neutralized. Biological oxidation was two-stage air activated sludge (AAS) and was seeded with a nearby coke liquor waste bio-solids. A fixed hydraulic detention time of 24 hours was used in 7-liter reactor-settler AAS units. Dilution was used to vary the food-to-micro-organism ratio, F/M. Bio-unit effluent was applied to 45 mm diameter char columns and char treatment resulted in 90 percent removal of TOC and 100 percent removal of color. TABLE 8-41 summarizes the char treatment for F/M ratios from 0.2 to 0.8 and shows that the maximum residual TOC was 35 mg/l. Although the

TABLE 8-38. SUMMARY OF UNSTRIPPED FOUL WATER CHARACTERISTICS, H-COAL (a)

Parameter	Value (b)
BOD (Total)	52,700
(Soluble)	51,200
COD (Total)	88,600
(Soluble)	88,000
TOD (Total)	13,200
Ammonia-Nitrogen	14,400
Nitrate-Nitrogen (c)	21,000
Phenol	6,800
Sulfide	29,300
Oil and Grease	608
Total Phosphorus	6.25
Suspended Solids	2
Volatile Suspended Solids	1
Total Dissolved Solids	5,300
Total Dissolved Fixed Solids	330
pH (pH units)	9.5
Chromium (Total)	0.10
(Hexavalent)	<0.01
Lead	<0.01
Nickel	0.06
Zinc	0.45
Cobalt	0.01
Copper	7.3

(a) From Item 8 in reference list.
(b) Values shown in mg/l, unless otherwise designated.
(c) Values reported are considered unrepresentative.

TABLE 8-38. SUMMARY OF UNSTRIPPED FOUL WATER CHARACTERISTICS, H-COAL (a)

Parameter	Value (b)
BOD (Total)	52,700
(Soluble)	51,200
COD (Total)	88,600
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TOD (Total)	13,200
Ammonia-Nitrogen	14,400
Nitrate-Nitrogen (c)	21,000
Phenol	6,800
Sulfide	29,300
Oil and Grease	608
Total Phosphorus	6.25
Suspended Solids	2
Volatile Suspended Solids	1
Total Dissolved Solids	5,300
Total Dissolved Fixed Solids	330
pH (pH units)	9.5
Chromium (Total)	0.10
(Hexavalent)	<0.01
Lead	<0.01
Nickel	0.06
Zinc	0.45
Cobalt	0.01
Copper	7.3

(a) From Item 8 in reference list.
(b) Values shown in mg/l, unless otherwise designated.
(c) Values reported are considered unrepresentative.

TABLE 8-39. ANALYSIS OF FOUL PROCESS CONDENSATE, SOLVENT REFINED COAL (a)

(mg/l unless noted)
Kentucky coal feed
Analyses by Water Purification Associates and Pittsburg & Midway

	pH=8.6	pH=8.2
Total Carbon	9,000	8,160
Total Organic Carbon	6,600	7,390
Inorganic Carbon	2,400 (b)	770 (b)
BOD (5 days)	32,500	
BOD (15 days)	34,500	
BOD (20 days)	>34,500	
COD	43,600	25,000-30,000
Phenol as C H OH	5,000	12,000
Total Kjeldahl N	8,300 (c)	15,000 (c)
Total Ammonia as N	7,900	14,000
Total Ammonia (meq/l)	465	824
Cyanide as CN	10	
Total Sulfur as S	10,500 (c)	16,200 (c)
Ca	0.47	
Mg	0.13	
Si	<0.5	

a. From item 7 of reference list
b. By difference
c. 22 analyses for N and S made between 10/5/75 and 12/9/75 were supplied by Pittsburg and Midway. Four of these analyses had extreme values and were arbitrarily eliminated. For the remaining 18 analyses the average total nitrogen was 12,600 mg/l with a standard deviation of 7,000 mg/l which is very random. The average ratio (moles NH_3)/(moles H_2S) was 2.0 with standard deviation of 0.17 which is quite reproducible

TABLE 8-40. SUMMARY OF TREATABILITY RESULTS, H-COAL WASTEWATERS, BIOLOGICAL OXIDATION*

	Period		
	A	B	C
Influent, mg/l			
BOD, Total	1,890	2,600	2,070
COD, Total	3,070	4,180	3,180
Phenol	750	1,450	760
Effluent, mg/l			
BOD, Total	26	36	24
BOD, Soluble	13	15	12
COD, Soluble	360	310	380
SS	60	40	50
Phenol	0.7	0.3	0.7
Parameters			
F/M, lb BOD/lb MLVSS-Day	0.06	0.17	0.22
Percent Removal			
BOD Basis	99.3	99.4	99.4
COD Basis	88.2	92.6	88.1

*From Item 8 in reference list

TABLE 8-41. TYPICAL ANALYSES OF CHAR-TREATED EFFLUENTS, SYNTHANE (a)

Reactor	TOC-to-Biomass Ratio	Influent (b)			Effluent			TOC Removal, Percent
		TOC, mg/l	pH	Color(c)	TOC, mg/l	pH(d)	Color(c)	
C-4	0.2	80	7.5	0.32	5	10.3	0.0	94
C-3	0.5	300	7.8	0.64	35	10.6	0.0	88
C-1	0.8	490	8.0	0.88	35	11.2	0.0	93
C-2	-	250	8.3	2.00	8	11.5	0.0	97

(a) From Item 2 in reference list

(b) Influent was effluent from biological unit

(c) Absorbance of color measured at 579nm

(d) Maximum pH value prior to color breakthrough ratio of COD/TOC is unknown for these waters, ratios for a few industrial wastewater effluents are reported (9) and range below 2.3:1. Thus, COD residual may be estimated at 80 mg/l. Figures 8-40(a) and 8-40(b) show color and TOC breakthrough curves respectively, for these studies.

The physical-chemical treatment of coke liquor is illustrated by a German study (10) that used isotherm, column, and pilot plant tests to design a demonstration moving bed Granular Activated Carbon plant for 25 m^3/hr (110 gpm) flow. Two years of pilot plant tests gave average removals of 99, 89.5, and 57.5 percent respectively for phenol, COD, and CN. Average feed content and removals for the pilot tests are given in TABLE 8-42. A simplified block flow diagram of the demonstration plant is shown in Figure 8-41. The plant design data are given in TABLE 8-43. The demonstration plant was designed for the maximum TOC removal (95 percent) obtained in the pilot plant. The authors include TABLE 8-44, an interesting comparison of TOC, COD, and BOD for coking plant effluent components. A fluidized bed thermal carbon regeneration system was used. The good results on phenol and COD removal make GAC a potential option to use in a physical-chemical treatment train between NH_3 recovery and tertiary treatment.

An unpublished study by AWARE (8) concerned treatment investigations and process design for the H-Coal liquefaction wastewater. This study included carbon isotherm tests for pretreated combined sour waters and for effluent from biological oxidation. Powdered Filtrasorb 400 activated carbon was used with 500 ml batch waste samples. COD and phenol removals were obtained and from these the Freundlich isotherm constants, K and 1/n, were evaluated. The results are summarized in TABLE 8-45. AWARE concluded that AC treatment was effective but would not be economically attractive because of the low equilibrum concentration (C_e) values required for discharge.

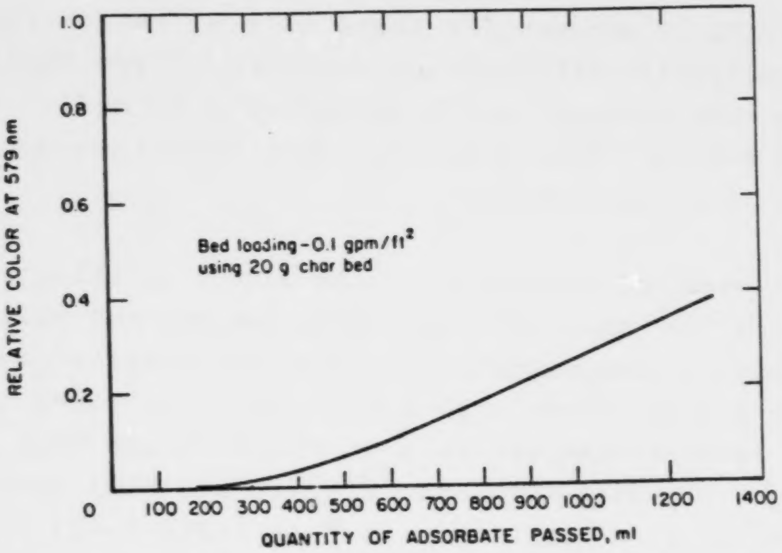

Figure 8-40(a). Color breakthrough with Montana char.*

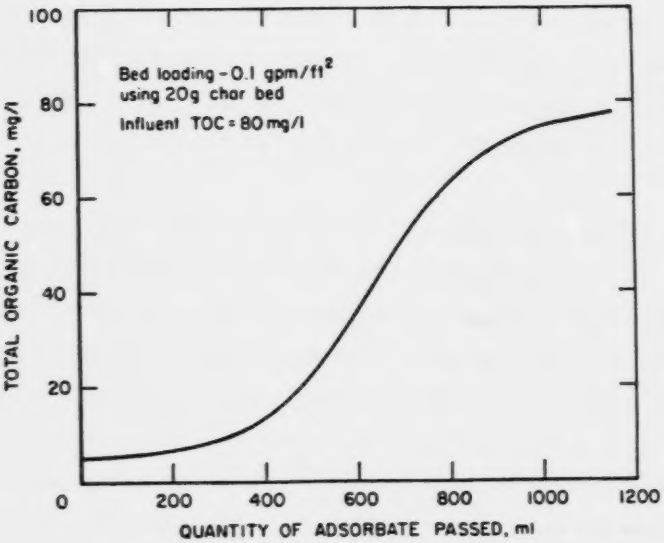

Figure 8-40(b). TOC breakthrough with Montana char.*

*From Item 9 in reference list

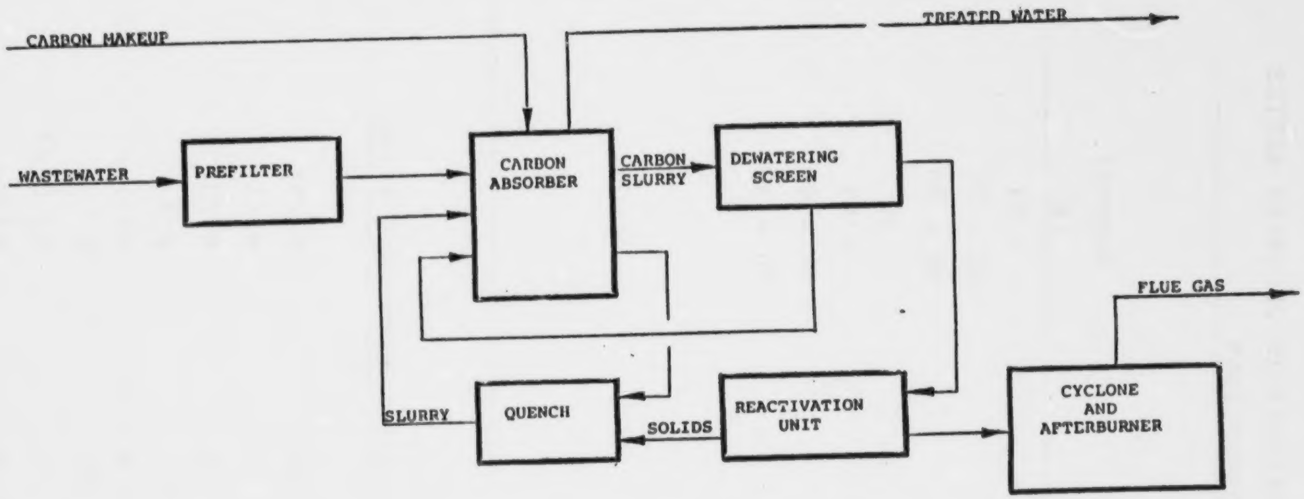
CARBON MAKEUP
TREATED WATER
WASTEWATER
PREFILTER
CARBON ABSORBER
CARBON SLURRY
DEWATERING SCREEN
FLUE GAS
SLURRY
QUENCH
SOLIDS
REACTIVATION UNIT
CYCLONE AND AFTERBURNER

Figure 8-41. Block flow diagram of demonstration plant, capacity 25 m^3 per hour.

(From Item 10 in reference list)

TABLE 8-42. CONTENT OF MAIN IMPURITIES IN DECANTER WASTES AND REMOVAL BY ADSORPTION*

	Content (mg/l)	Removal (%)
Phenols	650 - 1,400	> 99
Cyanide	5 - 35	45 - 70
Thiocyanate	120 - 450	30 - 80
Iron	40 - 150	30
Solids	300 - 3,000	> 99
TOC	800 - 2,000	85 - 95
COD	2,000 - 4,000	80 - 99

*From Item 10 in reference list

TABLE 8-43. DESIGN DATA FOR THE DEMONSTRATION PLANT

Waste Feed Rate	25	m^3/hr
Average TOC Content	1,000	mg/l
Flow Rate	10	m^3/hr
Activated Carbon: Bulk Density	450	g/l
Grain Size	2	mm
TOC Removal	95	%
Adsorbate Loading	70	kg C/m^3
Circulating Carbon	0.3	m^3/hr
Carbon Loss	0.14	kg/m^3

TABLE 8-44. CHARACTERISING VALUES OF COMPOUNDS FROM COKING PLANT EFFLUENTS

Compound in solution (100 mg/l)	TOC (mg/l)	COD (mg/l)	BOD_5 (mg/l)
NaCl	0	3.3	-
NH_3Cl	0	9.1	-
Na_2CO_3	11.3	1.8	-
NH_4SCN	15.8	64	-
H_2S	0	190	187
HCN	70.5	0.2	0
phenol	76.6	238	178
o-cresol	77.7	173	164
m-cresol	77.7	172	170
pyridine	75.7	0.6	115
benzene	92.5	0.8	0
naphthalene	93.6	23.4	0
anthracene	94.5	16.2	0
Cl-ion	0	5.5	-
SCN-ion	20.7	77.6	-
SO_3-ion	0	19.7	-

TABLE 8-45. H-COAL WASTEWATER TREATMENT: CARBON ISOTHERM CONSTANTS*

	1/n	K
Raw Wastewater		
COD Basis	0.51	1.53×10^{-2}
Phenol Basis	2.73	1.49×10^{-8}
Biologically Treated		
COD Basis	1.12	7.06×10^{-4}
Phenol Basis	3.41	7.60×10^{-3}

*Item 8 in reference list

References--

1. EPA 625/1-75-002a "Process Design Manual for Carbon Adsorption," October 1973, Chapters 2, 3. 871*

2. PERC/RI-77/13 "Treatability Studies of Condensate Water from Synthane Coal Gasification," November 1977. 797*

3. Ford, D.L., "Putting Activated Carbon in Perspective to 1983 Guidelines," Proceedings 1977 National Conference Treatment and Disposal Industrial Wastewaters and Residues, April 1977, Houston. 635*

4. Lawson, C.T., and J.A. Fisher, "Limitations of Activated Carbon Adsorption for Upgrading Petrochemical Effluents." AIChE Symposium Series: Water-1973, No. 136, Vol. 70. 866*

5. Hager, D. G., "Industrial Wastewater Treatment by Granular Activated Carbon," Industrial Water Engineering, Jan-Feb., 1974. 877*

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7. EPA 600/7-77-065, "Water Conservation and Pollution Control in Coal Conversion Processes." Water Purification Associates. 480*

8. Davis, G.M., Reap, E.J., and J. H. Koon, "Treatment Investigations and Process Design for the H-Coal Liquefaction Wastewater," Unpublished Report for Ashland Oil by AWARE, Inc., Dec. 1976. 678*.

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*Pullman Kellogg Reference File number

9. Davis, E.M., "BOD vs COD vs TOC vs TOD," Water and Wastes Engineering, Feb. 1971. 878*

10. Juntgen, H., and J. Klein, "Purification of Wastewater from Coking and Coal Gasification Plants Using Activated Carbon," Preprints 168th National Meeting ACS, Div. Fuel Chemistry, Vol. 19, No. 5, page 67. Sept. 1974. 875*

Chemical Oxidation of Effluents

Chemical oxidation is used primarily in disinfection of municipal water treatment effluents and in removal of specific industrial pollutants. Ozone, chlorine, chlorine dioxide, and hydrogen peroxide are the agents most applicable for chemical oxidation of wastewater effluents from coal conversion processes. Phenols, cyanides, ammonia, residual organics and some heavy metals are the specific substances to be treated by chemical oxidation of effluents.

There are hazards associated with storage, handling and use of the chemical oxidants:

- Ozone is an irritant and is toxic. Reference 2 discusses precautions for handling and use on pages 23 to 25 and 199 to 202. Because ozone is a more powerful oxidant than pure oxygen, the rate of combustion of oil, grease, and other organics in ozone ranges from extremely rapid to explosive.

- Chlorine is physically hazardous because it is stored and handled as a liquefied gas under pressure and is a toxic irritant. Its reactions with organics can be violent. Precautions for handling and use are described in detail by J. S. Sconce in "Chlorine: Its Manufacture, Properties and Uses," Reinhold Publishing Co., New York (1962). Safety considerations in its applications to wastewater are discussed in reference 1, pages 383 to 395.

- Chlorine dioxide is generated at the point of use by oxidation of sodium chlorite with chlorine. Because it is unstable, and explosive under certain conditions, it cannot be stored or shipped (1, p. 399). Both the reactants are hazardous and toxic as is the chlorine dioxide product.

- Concentrated hydrogen peroxide is a skin and eye irritant. It decomposes slowly by itself, and more rapidly in the presence of organics, some metals and light, into oxygen and water with the resulting potential of pressure buildup in its container to the point of rupture. As a powerful oxidant, its reactions with oxidizable materials may range from rapid to explosive.

Ozone--

Ozone (O_3) is a powerful, relatively expensive and unstable gaseous oxidant. The gas is generated onsite just prior to use in high capital cost equipment. Ozonation has been estimated (3) to add as much cost to tertiary effluent treatment as does activated carbon, but this appears to be questionable.

Ozone is produced economically only in 1 to 4 percent concentrations in air or pure oxygen and it is both unstable and only slowly reactive with some residual organics. For these reasons staged cocurrent contactors are required for efficient ozone utilization. Eductor-injectors were found (2, 3) to be efficient high shear mixing devices for ozone-wastewater contacting. Wynn, et al. (2), state that ozone transference is proportional to liquid flowrate at constant mixing efficiency, and the optimal gas to liquid volumetric ratio was nearly 1.0 for the 55 gpm pilot flow used. Reference 2 discusses mixing and ozone transference on pages 109 to 123 and economic analysis for full-scale ozone treatment on pages 127 to 170.

In a study of ozonation treatment of hospital effluents (4) a synthesized permeate from reverse osmosis (RO) containing five known organics, methyl, ethyl,and isopropyl alcohols were found to be sustantially reduced but acetone and acetic acid were not: acetic acid concentrations increased during ozonation and acetaldehyde appeared.

In another study (5, 6) the secondary effluent from an air activated sludge (AAS) treatment was passed through a membrane filter having 0.45 micrometer pores. The filtrate was then subjected to ultrafiltration (UF) where organics having molecular weights greater than 18,000 (humic acids) were retained. The ultrafiltration permeate was subjected to reverse osmosis (RO) to yield a retentate fraction containing organics with molecular weights ranging from 150 to 18,000 (fulvic acids) and a permeate fraction containing organics with molecular weights less than 150. The secondary effluent from AAS, the UF retentate and the RO retentate and permeate were all subjected to ozonation with the following results:

- Permeate analyses varied when the parameter F/M in the AAS system was varied.
- Variation in ozonation results reflected the variation in the F/M parameter.
- Ozonation of the AAS effluent feed showed a high ozone demand even without TOC reduction.
- Acetone and acetic acid were destroyed by ozonation.
- Acidic pH enhanced the ozonation reactions.
- Organics in the RO retentate, with molecular weights ranging from 150 to 18,000, exhibited a greater ozone demand than did those in the UF retentate with molecular weights above 18,000 or those in the RO permeate with molecular weights below 150.

Chlorination of the UF and RO retentates produced volatile halogenated organics, whereas ozonation did not. Ultraviolet irradiation of the permeates prior to ozonation enhanced the removal of organics.

Research is needed to determine the changes in chemical species produced in AAS effluents with changing F/M ratios and also the effects of ozonation of these compounds.

Ozone has been used for reduction of foaming, color, turbidity, odor, and heavy metals (3, 5, 7, 9) as well as bacteria and viruses. Limited data (8) indicate 10 to 30 percent BOD/COD reductions at the 10 to 25 mg/l ozone dosage used to achieve disinfection, and little ammonia removal. Diaper (8) states that a typical industrial cyanide application destroyed 15 mg/l cyanide with 80 to 88 mg/l ozone. An oil refinery used 20 to 40 mg/l ozone to reduce effluent phenols to less than 0.015 mg/l. Diaper found 1973 production costs were 8.0 and 3.5 cents per pound for ozone from air and oxygen, respectively, apparently exclusive of amortization costs, and capital costs of $500 to $1,000 per pound per day capacity. Single units of 1,000 lb/day capacity are available.

A recent paper by Hardisty and Rosen (9) gives some details on ozonating strong industrial wastes for removal of cyanides, phenol and COD. Generally non-ferrous complexed cyanides, simple cyanides and cyanates were readily destroyed with 1.85 to 3.8 mg ozone per mg CN. Ferrous complexed cyanides were reduced but not totally destroyed with 10 mg ozone per mg total CN and residuals were ≤ 2 mg/l total CN . Complete oxidation of phenol to CO_2 and H_2O would require 14 moles ozone per mole phenol removed, but partial oxidation may be all that is needed. At the <1 mg/l phenol concentrations in biological oxidation effluents the intermediate organics formed may not be troublesome. COD reductions were given for raw wastewaters (9) and showed generally higher removals at higher ozone doses. Petrochemical wastes required 2.5 mg ozone/per mg COD removed to reduce COD from 2,400 to 500 mg/l.

Ozone is more stable at acid pH than at basic (5, 10), but the optimum ozone treatment pH will depend on the waste and the desired results. Apparently acid pH is also good for disinfection (5) but basic pH will enhance ammonia removal (8, 10) and organics reduction (5) including phenols (9). Ozone treatment of coal conversion wastewaters should be studied to determine applicability and optimization for metals removal (7), cyanide, ammonia, and COD reduction (5, 8, 9) and enhancement of follow-on activated carbon treatment (11).

Chlorine--

Chlorination with elemental chlorine, chlorine dioxide or various chlorinating compounds such as sodium and calcium hypochlorite is the most common chemical oxidation method practiced for wastewater treatment at the present time. Nonetheless, chlorination applications to coal conversion wastewaters must be limited to intermediate treatment steps to avoid formation of trihalomethanes and chlorinated phenols in final effluents. Cyanide and ammonia destruction are two possible applications. Pretreatment for ammonia removal after ozonation (12) and before activated carbon adsorption (13) is another possible application.

Alkaline chlorination of cyanide wastes, developed for treatment of concentrated metal plating wastes (14), is flexible and capable of treating dilute or concentrated wastes to cyanide residuals of 0.2 mg/l. In practice, about 8 parts chlorine by weight plus 7.3 parts NaOH are needed per part cyanide destroyed.

In a discussion of "breakpoint" chlorination for ammonia removal, it is shown that ammonia can be oxidized to nitrogen gas with about 8 parts chlorine per part ammonia destroyed (13). Sodium hypochlorite (NaOCl) is recommended instead of chlorine for ammonia oxidation because local pH depression, with its attendant problems, does not occur and less TDS will be added if neutralization is required. Chlorine will destroy alkalinity at 14.3

mg/l as $CaCO_3$ per mg/l ammonia oxidized. If the wastewater alkalinity is low, pH control will be required using lime or NaOH. Thus chlorine plus lime will add 12.2 mg/l, and NaOCl only 7.1 mg/l, of TDS per mg/l ammonia oxidized.

Hydrogen Peroxide--

Hydrogen peroxide, H_2O_2, is bulk transported as 70 percent solution in water and as 35 percent and 50 percent solution in drums. The concentrated solution is diluted as needed at the point of use. Doses and usage are based on 100 percent H_2O_2. The major wastewater uses of hydrogen peroxide are for destruction of sulfide odors, correction of AAS sludge "bulking", phenol destruction and as a supplemental oxygen supply for overloads or equipment breakdowns. Its advantages over chlorination include oxidation without TDS increase. Acid to neutral pH's allow the oxidation of H_2S to elemental S. Refining uses described (15) included sulfide oxidation in sour waters during sour water stripper maintenance and supplemental oxidation of phenols and of thiosulfate effluents before discharge to a municipal sewer (16).

The literature describes several applications of hydrogen peroxide to the problems of rising or bulking sludge. Rising sludges occur whenever nitrates present are reduced to elemental nitrogen at a rate sufficient for the entrapped nitrogen to float the sludge. Ironically, underloading AAS plants often results in increased nitrification of ammonia, which is desirable per se, and thus increases sludge rising potential. Bulking sludges are those that will not settle due to their density and flocculant nature. The primary cause of bulking is low oxygen in the aeration tanks. Hydrogen peroxide has been used in the secondary clarifier at about 7 mg/l for rising sludge and in the return sludge lines at 20 to 60 mg/l for bulking and sludge odors (16) as well as a means for increasing the dissolved oxygen in aeration tanks.

Application Summary, Doses and Costs--

Ozone may be used at 20 to 40 mg/l for residual phenol reduction and at 2 to 4 mg ozone per mg cyanide destroyed to achieve cyanide residuals of $\leq$ 2 mg/l total CN . COD reduction in raw petrochemicals wastes required 2.5 mg ozone per mg COD destroyed. Ozone costs are estimated to be as follows in tertiary treatment for a COD reduction from 35 to 15 mg/l, assuming 80 percent utilization (3) of ozone produced from oxygen.

Capacity, MGD	Capital Cost*	Operating Cost,** ¢/1000 Gal
1	$ 556,000	54.4
10	3,025,000	31.6
100	20,900,000	22.4

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* Capital cost (3) for 1969 updated to 1977 by use of 2.8 multiplying factor.

** Operating cost (3) for 1969 updated to 1977 by use of 2.0 multiplying factor.

These estimates agree reasonably well with Diaper (8), whose 1973 rough estimate is $500 capital per pound ozone per day capacity for an oxygen-fed plant. The 20 mg/l COD reduction would require about 520 pounds ozone/day at 80 percent utilization per MGD plant flow.

Chlorine could be used in intermediate steps to reduce ammonia and cyanide residuals. Either use requires about 8 mg/l chlorine per mg/l ammonia or cyanide destroyed. Cyanide residuals of 0.2 mg/l are possible. Chlorine is currently (1978) $84-115/ton in tank cars.

Hydrogen peroxide could be used for biological waste treatment plant sludge control at 20 to 60 mg/l in the return sludge, for odor control and to supply supplemental oxygen. Current bulk costs are 40 to 44 ¢/lb (100% H_2O_2 basis) at 70 to 35 percent by weight concentrations.

References for Chemical Oxidation of Effluents--

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2. EPA-R2-73-146, "Pilot Plant for Tertiary Treatment of Wastewater With Ozone," Jan. 1973. 961*

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4. Elia, V. J., Clark, C. S., McGinnis, K. T. and R. N. Kinman, "An Examination of Products Formed by Ozonation in a Wastewater Reuse System." Abstract of paper given at 49th Annual Conference, Water Pollution Control Federation, Oct. 1976, Minneapolis, Minn.

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15. Strunk, W. G., "Hydrogen Peroxide for Industrial Wastewater Pollution Control," Proceedings 1977 Conference on Treatment and Disposal of Industrial Wastewaters and Residues, pp. 119-125, April 1977, Houston. 635*

16. Ochs, L. D. and C. W. Cooke, "Industrial Applications of Hydrogen Peroxide" in Water - 1975, AIChE, 151, (71) 1975, pp. 59-63, edited by Bennett. 518*

COSTS OF WATER TREATMENT

Basis for Development of Costs

As the quotations of costs in the preceding presentations on individual water treating methods indicate, there are many references to costs in the open literature. These data require updating of investment and operating costs and correlation in order to obtain a proper perspective. Some of the short references indicate a rather wide variation for the same process, depending upon the feed stream composition and the residual of contaminants in the effluent, and all data that could be desired are rarely present. In some cases the effective date of the cost data is not specified, or the split between amortization and operating costs is not clear, or operating costs are based on outdated figures and cannot be updated due to lack of detail or the influent and effluent compositions are either not specified or apply to wastes far different from those in coal conversion plants. Most of these references are to municipal plants.

We have approached the problem through correspondence with process licensors or equipment vendors and have requested budget cost figures based on specific cases in which all available data on influent quantity and composition are supplied together with the required or desired residuals of contaminants. In order to supply such data quantities from three representative conceptual designs have been used. These are:

- Gasification with p/o/t production as exemplified by the Lurgi process in the C.F. Braun design for western coal (294, 295, 296).
- Gasification without p/o/t production, of which the Bi-Gas process is an example, described in the C.F. Braun design

for western coal (294, 295, 296).

- Liquefaction, as described in the SRC-II process design by R.M. Parsons for eastern coal (814).

Compositions of wastewater used with the quantities from the above designs were those from other sources, as described previously in the section on "Analyses of Waste Streams." For gasification without p/o/t production, Koppers-Totzek analyses were substituted, since no Bi-Gas data have been published. For liquefaction, H-Coal data were more complete and these were judged similar to those which would be obtained from SRC-II.

In order to develop the compositions of streams intermediate in the treating sequence, it was necessary to make assumptions, based on best engineering judgement, of the efficiency of the treating steps which preceded the step for which costs were requested. Obviously, this procedure must eventually be confirmed by actual testing on wastewater from bench scale or pilot plant operation or in a demonstration plant. All licensors and vendors quite reasonably state that data supplied must be verified in this manner.

Equipment vendors usually supply only the cost of unassembled equipment, f.o.b. their supply point. Costs of freight, assembly and installation have been estimated by Pullman Kellogg, using factors which we specify. These are judgement factors that are based on experience and on opinions of qualified personnel.

Licensors usually supplied total capital cost directly, based on actual installations. These included Zimpro, Inc. (wet air oxidation and biological systems employing powdered carbon regenerated by wet air oxidation), Chevron Research Co. (stripping and ammonia recovery), U. S. Steel (stripping and ammonia recovery) and American Lurgi (Phenosolvan phenol recovery).

Licensors furnish process requirements of chemicals, steam and electric power from which operating costs based on up-to-date utility values can be assembled. Equipment vendors supply only spotty data on utility requirements or none at all (e.g., no pumping costs). Time and manpower remaining in our project did not permit detailed estimation of all these costs, so where necessary we have resorted to the use of estimated factors.

In order to update costs reported for prior years we have obtained cost indices from Pullman Kellogg (process plants) and Engineering News Record (ENR), the Water Quality Office-Sewage Treatment Plant Index (WQO-STP), and compilations by EPA of the Large City Advanced Treatment (LCAT) and the Small City Conventional Treatment (SCCT) indices.

The Pullman Kellogg cost index is shown in Figure 8-42. and illustrates the sharp rise in direct materials costs. Direct materials, as defined by Pullman Kellogg, includes all furnaces, exchangers, converters, towers, drums and tanks, pumps and compressors, special equipment, utility conveying and safety equipment, site preparation and foundations, structural steel, buildings, piping, electrical, instruments, paint, insulation, and catalysts. To the direct materials costs are added operations costs, which include construction labor and supervision, design engineering, freight, insurance, and contractor's overhead and profit. The sum of direct materials and operations costs is the total "as built" cost of the plant. An average split of these two categories for process type plants (ammonia, ethylene, phenol, and crude oil distillation) is 60 percent direct materials and 40 percent operations costs. In Figure 8-42 a line for construction labor is a representation of the largest component of operations cost and it can be seen that this has risen at a lower rate than direct materials in the 1970-1977 period. Projections included beyond this period are speculative but are

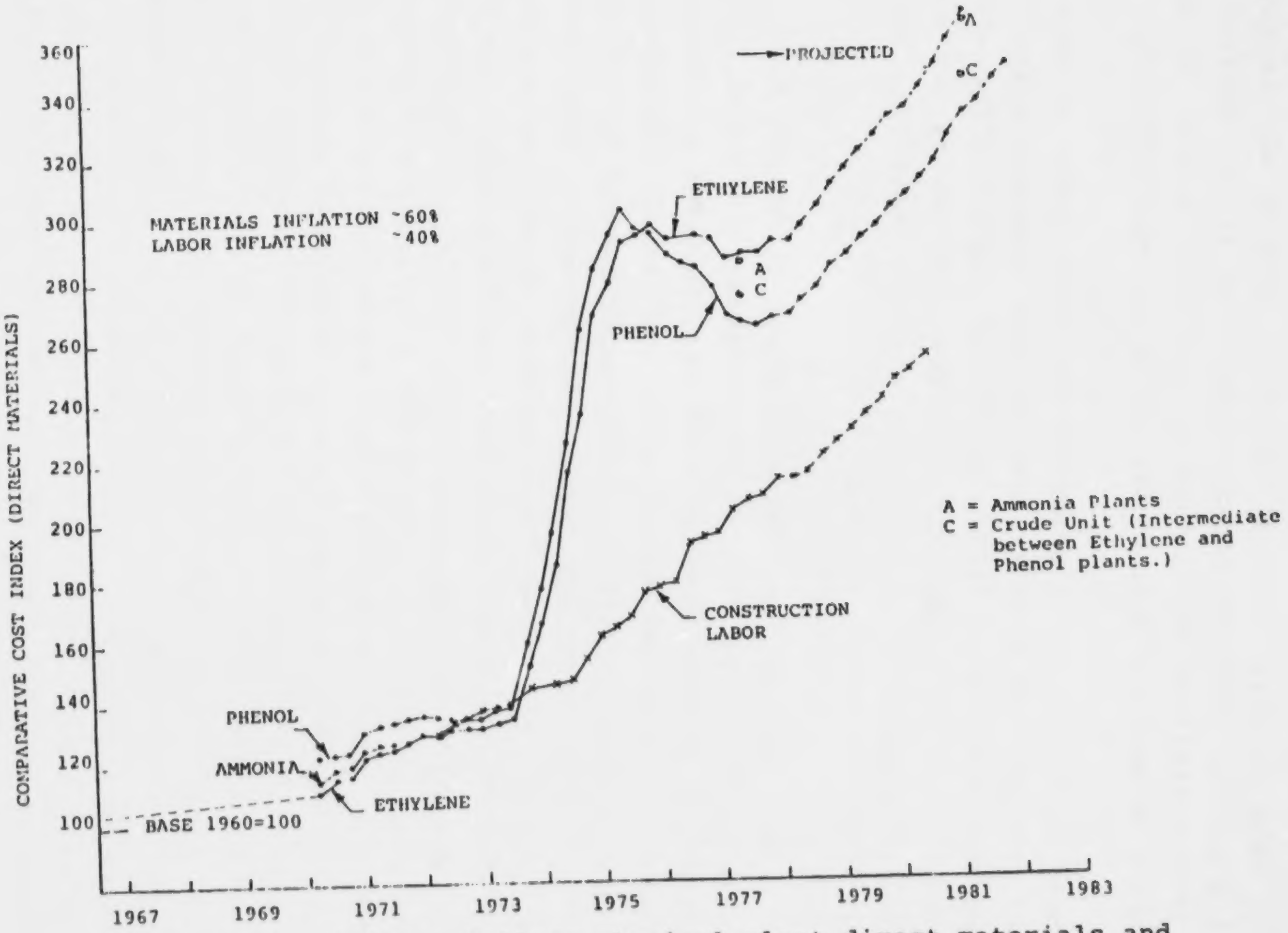

Figure 8-42. Pullman Kellogg chemical plant direct materials and construction labor costs.

estimated at about 7.5 to 7.8 percent per year through 1980, with labor costs at the same escalation rate. The Engineering News Record (ENR) index is shown for comparison in Figure 8-43. This is for skilled labor and is generally used more for estimating operating labor costs than for estimating construction labor costs.

Indices specifically for water treatment plants have been compiled by EPA and its predecessor agencies and by the Federal Water Pollution Control Federation (FWPCF). These are for municipal water treatment plants, not industrial water treating. Some of these indices are plotted on Figure 8-44 to illustrate increases of costs over the period 1970-77 for comparison with the Pullman Kellogg process plant index. One illustration of comparison is as follows:

	SCCT	LCAT (EPA)	Pullman Kellogg		
	(EPA)	Composite	Equipment	Labor	Composite
Factor for 1973	190	100	132	132	132
Factor for 1977	273	138	265	212	244
Percent increase	44	38	100	61	85

The inference appears to be that water treatment plant costs have not risen as fast as process plant costs. Although this may be true for municipal water treatment, it must be remembered that industrial water treatment is more complex than municipal water treatment, is likely to be more costly than municipal treatment and is likely to be less regulated. We feel that the increase probably lies between the two types of plants compared above, say 50 to 60 percent in the last four years.

Supporting cost data from open literature are those from Water Purification Associates (480, 612), AWARE, Inc. (643) and Bechtel (WateReuse-1975-AIChE).

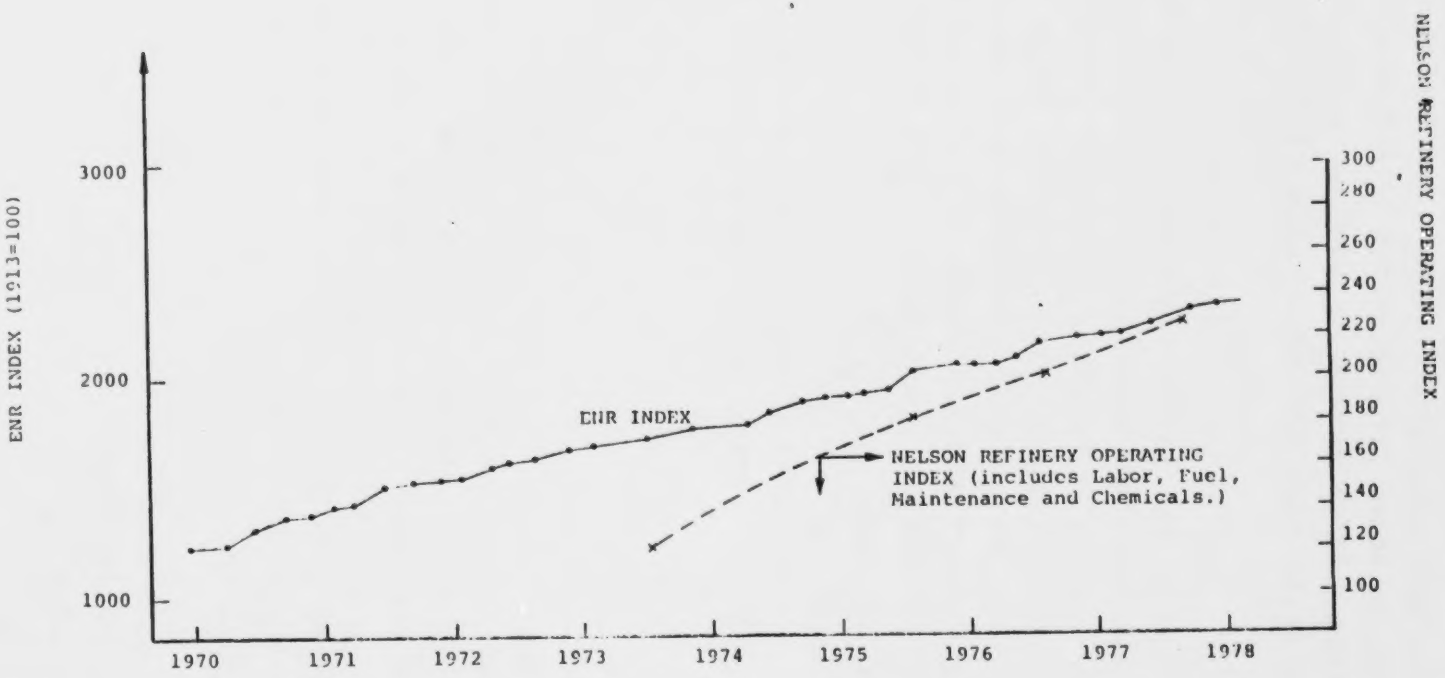

Figure 8-43. Engineering News Record (ENR) skilled labor index.

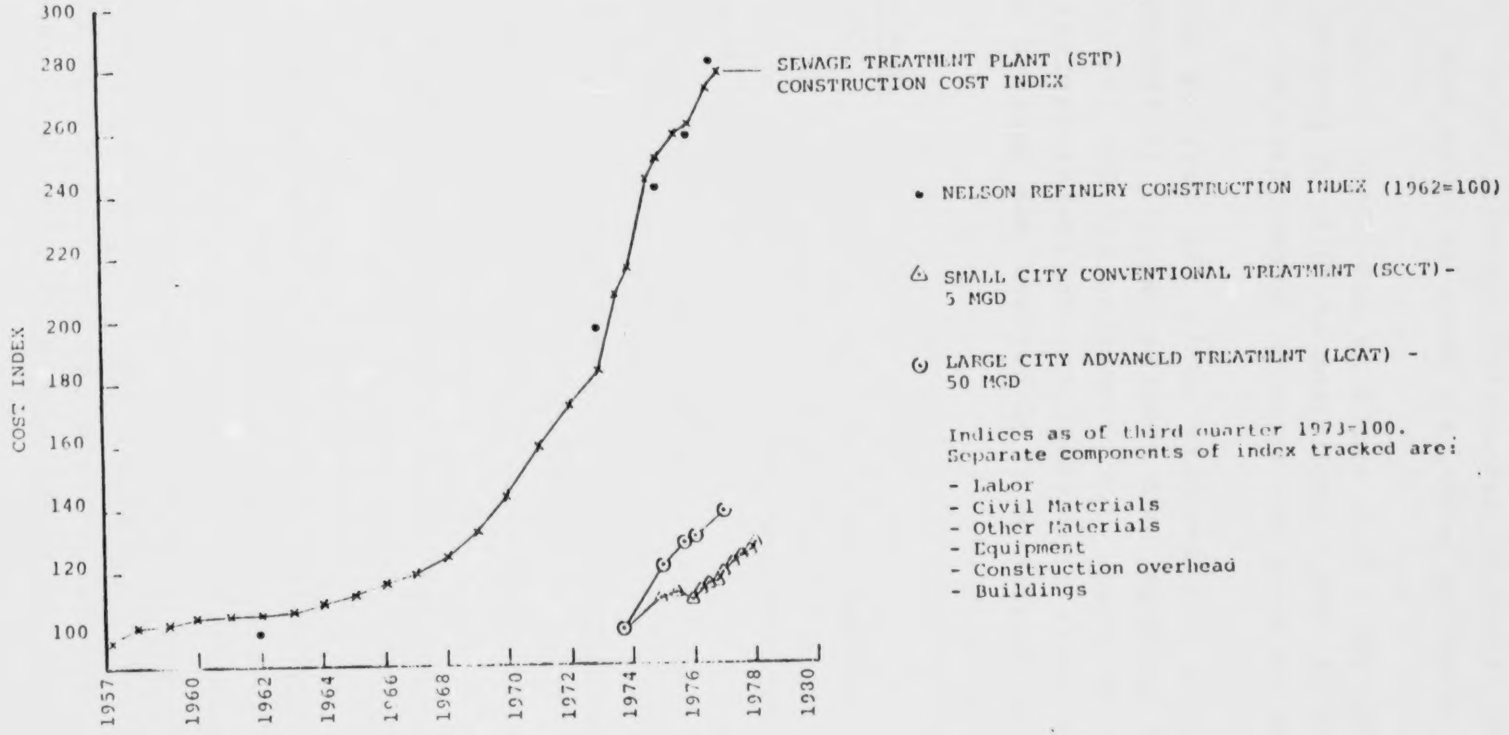

Figure 8-44. Cost indices maintained by EPA.

From the Water Purification Associates reports, water treating costs, in rough ascending order, are shown in Table 8-46.

The year basis for the W.P.A. costs is not clear in all cases, but it appears to be 1975. Escalation of 10 to 15 percent should probably be added to bring the costs to current levels.

Bechtel presented comparative costs for several types of evaporators, ion exchange, and reverse osmosis in "WateReuse - 1975," published by AIChE. For processing 810,000 lbs/hr (2.33 MGD*) of cooling tower blow-down, the following capital costs were extracted (an escalation of 30 percent was applied to make the costs more nearly current):

	Cost - $MM
Three evaporator-crystallizers	20.4
Two multi-stage flash (540,000 lb/hr) plus Two evaporator/crystallizers (270,000 lb/hr)	11.83
Two multi-stage flash units (700,000 lb/hr) plus One evaporator-crystallizer (100,000 lb/hr)	9.36
Four ion exchange units plus reverse osmosis plus Two multi-stage flash (326,000 lb/hr) plus One evaporator-crystallizer (24,000 lb/hr)	6.76

Comparison of the Bechtel figures with the previous tabulation of Water Purification Associates figures leads to the conclusions that:

- o W.P.A. figures are probably low for evaporation.
- o Combinations of reverse osmosis and ion exchange with small evaporators for their residues is significantly cheaper than evaporator/crystallizers alone. "Capitalized" operating costs for one year were also included in this article and these supported the conclusion that the fourth case above is to be preferred.

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*Million gallons per day

TABLE 8-46. WATER TREATING COSTS FROM WATER PURIFICATION ASSOCIATES REPORTS

Process	Feed, Rate, Investment	Operating & Amortization Cost/ 1,000 gals
Electrodialysis	Cooling Tower Blowdown 1 MGD = $ 600,000 10 MGD = $3,200,000	$0.56 (N. Mex.)
Reverse Osmosis	1 MGD = $1,700,000 10 MGD = $10,700,000	0.73 0.43
Ion Exchange	Boiler Feed Water 2.5 MGD = $3,200,000	>$0.92 (Energy requirements not included)
Bioxidation (Trickling Filter plus High Purity Oxygen Activated Sludge)	Wastewater (COD = 13,700 mg/l) 3.0 MGD* = $8,700,000 (incl. equalization, sludge thickening & filtration. Phenols < 1 ppm)	$3.10 (Most cases of bio-oxidation would be lower. Loading selected is extremely high)
Activated Carbon	Wastewater (COD = 2,000) 3.0 MGD = $4,200,000	$2.00
	Wastewater (COD = 10,000) 3.0 MGD = $11,400,000	$9.00
Phenol Extraction	Wastewater (6,000 ppm phenols) 2.9 MGD = $9,000,000 (94% phenol recovery)	Gross = $3.57 Net = $2.57 (with phenol sale at 2 ¢ per lb)
Ammonia Recovery	Wastewater (6,000 mg/l ammonia) 3.6 MGD = $8,000,000 (200 ppm NH residual, 95% NH recovery)	Gross = $4.03 Net = $0.73 (with ammonia sale at $140 per ton)
Evaporation	Wastewater 3.0 MGD = $6,000.000 (95% recovery)	$3.50-4.00
Wet Oxidation	Wastewater (95% COD removal) 3.0 MGD = $22,000,000	$5.57 (no credit for steam or power generation)

This same Bechtel article also gave the following breakdown of cost which may have general application:

Item	Percent of Purchased Equipment
Installation	35-45
Instrumentation	13
Piping	31
Electrical	10-15
Buildings	29
Total add-ons to obtain capital cost	118-133

AWARE, Inc. (Associated Water and Air Resources, Inc., Nashville, Tenn.) present a course on water treating (643) in which an equation for capital cost of activated sludge in municipal water treatment is shown as:

$$\text{Capital Cost} = 226\ (\text{MGD}) + 45.6\ (\text{MGD})^{0.18}$$

With this equation, the capital cost (1969) for a 3.0 MGD plant is $733,700. When the Pullman Kellogg escalation factor of 2.52 is applied for updating, the capital cost (1977) becomes $1,849,000.

It may be unfair to compare this developed capital cost with the $8,000,000 cost developed by Water Purification Associates for activated sludge without equalization and sludge handling, since W.P.A.'s influent was much higher in COD and their effluent was much purer than municipal practice.

For industrial waste capital costs AWARE presents the equation:

$$\text{Cost} = Q^{w}(17(S_o/S_e)^{0.77}+215)(1.05+(0.044)/kX_v)$$

where $w = 0.69 + 0.00019S_o$

S_o = Influent BOD, mg/l

S_e = Effluent BOD, mg/l

Q = Flow rate, MGD

k = Reaction rate, l/mg/hr

X_v = MLVSS,* mg/l

Applying the quantities and rates from the Water Purification Associates example:

$S_e = 1800$

$S_o = 18{,}000$

$Q = 3.0$

$k = 0.000833$

$X_v = 4000$

yields calculated capital cost of $30,623,000.

The installed cost calculated by the AWARE equation appears to be excessively high, particularly in contrast to costs developed by Braun for western coal (294,295,296) which in our opinion generally appear to be on the high side. Braun, for example, shows the equipment cost at $3,200,000 and installed cost at $19,000,000 for an activated sludge system fed with Lurgi wastewaters from which phenol had previously been extracted.

We conclude that either the AWARE, Inc. equations for activated sludge units do not apply to the cases under consideration, or that the reaction rate is incorrect, since the equation for municipal plants apparently yields low capital costs and the equation for industrial activated sludge yields very high capital costs when the quantities and compositions employed in the Water Purification Associates report, with no preceding phenol extraction, are used.

*Mixed Liquor Volatile Suspended Solids

Other cost equations given by AWARE, Inc. are as follows for domestic and combined wastes:

Treatment Unit	1969 Capital Cost ($1,000)
Pretreatment	$19\ (Q)^{0.63}$
Primary Sedimentation	$17.3(SA) + 6.7\ (SA)^{0.1}$
Blower House	$13.6 + 7.6\ (CFM/1{,}000)$
Sludge Return Pumps	$4.6 + 1.45\ Q$
Final Sedimentation	$16.2(SA) + 6.9/(SA)^{1.3}$
Chlorination	$11.6\ (Q)^{0.47}$
Anaerobic Digester	$V(1.34 + 13.8/V^{0.87})$
Vacuum Filters	16.5 + 48 (Area in sq.ft/100)
Sludge Incinerator	$(S/24000)(170 + 7.15\ S^{0.61})$
Control House	$51.6\ (Q)^{0.7}$
Tertiary treatment (lime clarification, mixed media filtration)	$\text{Cost} = 200\ (Q)^{0.76}$
Carbon adsorption	$\text{Cost} = 202\ (Q)^{0.86}$

where SA = surface area in 1,000 sq. ft.
V = volume in 1,000 cu. ft.
S = sludge production, lb/day
Q = flow rate, million gallons/day (MGD)

Formulas are also presented for annual operating costs in $/MGD and curves for capital and operating costs are presented for equalization, oil separation, neutralization (which includes equalization, oil skimming, chemical addition and flocculation, sedimentation, and vacuum filtration of sludge), primary classification, aerated lagoons, chemicals coagulation, dissolved air flotation, sand/mixed media filtration, chrome reduction, and cyanide destruction. All these cost equations and curves must be converted from 1969 cost basis to current cost basis using the WQO-STP index for capital cost and ENR index for operating cost.

It should be noted that EPA changed from use of WQO-STP to other indices (LCAT, SCCT, and CUSS) in 1973. Only the SCCT is considered applicable to the project. In general, this source of cost information was used only to supplement budget costs acquired from vendors and licensors.

In order to convert purchased equipment costs from vendors into installed cost, we compared the factors reported earlier by Bechtel with a plant constructed in South America by Pullman Kellogg. This plant was designed by Engineering-Science, Inc. of Texas and costs on a Gulf Coast basis were cited. Also, the new SCCT index adopted by EPA has a breakdown which is included for comparison:

	Percent of Purchased Equipment		
	Engineering-Sciences, Inc.	Bechtel	SCCT (EPA)
Installation	306.8*	35-45	252.5**
Labor	Included	-	233.0
Instrumentation	14	13	Apparently included in installation and purchased equipment
Piping	22	31	
Electrical	17	10-15	
Buildings	1.6	29	135.9
Buildings, including paving	17.9	-	-
Purchased equipment	100.0*	100.0	100.0

*Purchased equipment in this case includes some installation costs

**Civil and other material in installation

It is evident that there is a wide variation in the factors that could be employed. Part of this variation can be explained by defining the differences in the installations:

- The Engineering-Sciences, Inc. column represents an impoundment basin of 13 million gallons, an equalization basin of 3.3 million gallons, an activated sludge system to process 18.3 MGD (six 2,300,000 gallon basins) and sludge handling consisting of aerobic digestion and thickening. It should be noted that a labor component is included for installation of "purchased equipment" for this column. Therefore all other component percentages are low.
- The Bechtel column represents such things as ion exchange, reverse osmosis, and evaporators where purchased packaged equipment is the main cost and little construction labor is required for installation (minimum of earthwork, foundations, and piping).
- The EPA Small City Conventional Treatment (SCCT) column is closely similar to the first column. It represents a 5 MGD municipal plant containing bar screen, grit chamber, primary clarification, conventional activated sludge, chlorination, gravity thickening, and vacuum filtration. It indicates that purchased equipment could be only 11.3 to 20.3 percent of the total cost (average is 14.05 percent), depending on location in the United States.

Development of Costs for Activated Sludge Systems

Envirotech Equipment Co. (Division of EIMCO) supplied equipment costs for four hypothetical cases for which we provided flows and compositions. The cases are indentified as follows:

Case No. I: 4.032 MGD of wastewater from Lurgi gasification (Braun design)

Case No. II: 0.533 MGD of wastewater from liquefaction (Parsons design)

Case No. III: 0.85 MGD of average wastewater from p/o/t-producing gasification

Case No. IV: 0.80 MGD of wastewater from liquefaction (W.P.A. basis)

TABLE 8-47 shows the size and cost of the equipment to be supplied. Flash mixing, flocculation, and dissolved air flotation precede the 2-stage activated sludge biological oxidation system. Sludge is thickened by dissolved air flotation and aerobic digestion. Biological clarified effluent is filtered. Total equipment costs are:

Case I	$1,120,000
Case II	$ 764,000
Case III	$ 467,000
Case IV	$1,288,000

In order to arrive at installed cost many other items need to be estimated:

- Freight
- Installation labor
- Aerator platforms
- Aerator basins
- Surface preparation and painting
- Site preparation and drainage
- Pumps
- Instruments
- Piping
- Electrical equipment
- Laboratory equipment
- Foundations
- Buildings
- Engineering costs
- Administrative costs
- Contingency
- Contractor overhead and profit

The Braun report on Western coal that was cited previously shows a factor of 5.94 installed cost to equipment cost ratio for

TABLE B-47. PULLMAN KELLOGG COAL CONVERSION STUDY
(ENVIROTECH)

	CASE I GASIFICATION	CASE II LIQUEFACTION	CASE III GASIFICATION	CASE IV LIQUEFACTION
FLOW (GPM)	2800	370	590	555
FLASH MIX - SIZE, COST	17' Ø x 8' SD $13,000	7' Ø x 7' SD $8,000	9' Ø x 7' SD $9,000	9' Ø x 7' SD $9,000
FLOCCULATOR - SIZE, COST	38' Ø x 8' SD $24,000	5' Ø x 7' SD $9,000	21' Ø x 7' SD $10,000	(1) 21' Ø x 7' SD $10,000
DAF-BR - SIZE, COST	60' Ø x 8' SD $105,000	25 Ø x 7' SD $52,000	30' Ø x 7' SD $64,000	(1) 30' Ø x 7' SD $64,000
CLARIFIER-C2S - SIZE, COST 1st Stage	130' Ø x 12' SD $85,000	50' Ø x 12' SD $48,000	65' Ø x 12' SD $53,000	(1) 65' Ø x 12' SD $53,000
AERATION - SIZE, COST 1st Stage	(8) 150 HP $275,000	(8) 150 HP $275,000	(2) 125 HP $57,000	(16) 150 HP $549,000
CLARIFIER-C2S - SIZE, COST 2nd Stage	130'Ø x 12 SD $85,000	50' Ø x 12' SD $48,000	65' Ø x 12' SD $53,000	65' Ø x 12' SD $53,000
AERATION - SIZE, COST 2nd Stage	(2) 100 HP $53,000	(2) 50 HP $34,000	(2) 25 HP $25,000	(2) 125 HP $57,000
DAF SLUDGE THICKENER - SIZE, COST	30' Ø x 7' SD $64,000	30' Ø x 7' SD $64,000	25' Ø x 7' SD $55,000	45' Ø x 8' SD $80,000
DIGESTION SUBMERGED TURBINE AERATOR - SIZE, COST	(3) 150 HP $121,000	(3) 150 HP $121,000	(2) 40 HP $42,000	(6) 150 HP $233,000
BLOWER - SIZE, COST	(3) 75 HP $45,000	(3) 75 HP $45,000	(2) 25 HP $18,000	(6) 75 HP $90,000
TERTIARY FILTERS - SVG SIZE, COST	(3) 22' Ø $250,000	(2) 10' Ø $60,000	(3) 10' Ø $90,000	(3) 10' Ø $90,000
TOTAL EQUIPMENT	$1,120,000	$764,000	$476,000	$1,288,000

biological oxidation for Lurgi and 5.2 for HyGas. For ammonia recovery and stripping for Bi-Gas, the ratio is 4.4.

Volume and depth of the aeration basins was supplied as shown below. These will be assumed rectangular, constructed of one foot thick reinforced concrete, with 3 to 4 feet projecting above grade and the rest below grade. Costs of these basins are not included in equipment costs.

Case No.	I	II	III	IV
1st stage basin				
Volume, CF	1,340,000	1,390,000	303,000	2,600,000
Depth, ft.	20	20	15	20
2nd stage basin				
Volume, CF	253,000	132,000	58,000	245,000
Depth, ft.	15	12	9	15
Sludge Digester Basin				
Volume, CF	150,000	160,000	78,000	303,000
Depth, ft.	20	20	15	20

Because of the shortage of time and manpower it was not possible to define the system in detail. We therefore resorted to factors to obtain the investment cost to equipment cost ratio.

One approach is to use the EPA/SCCT breakdown for municipal plants where equipment cost averages 14.05 percent of total cost for the U. S. Investments obtained in this manner would be:

Case I	$7,971,530
Case II	$5,437,720
Case III	$3,387,900
Case IV	$9,167,260

The Water Purification Associates report (480) on page 374 shows the following basis for obtaining capital costs of air activated sludge plants:

Aeration Basin	(Configuration as illustrated on p. 372)	$240 per cu yd.
Aerators		$600 per installed HP
Clarifiers	(Curve - p. 301)	
DAF thickeners	(Curve - p. 375)	
Vacuum filters	(Curve - p. 375)	

Operating Costs

Amortization	15 percent of capital/yr.
Maintenance	3 percent of capital/yr.
Electricity	2¢/KWH, 8,000 hr/yr.
Phosphorus	42¢/lb as P

Applying this implied method the following is obtained:

	Capital Costs, Millions of Dollars			
	Case I	Case II	Case III	Case IV
Aeration basins	2.98	2.79	0.73	5.20
Aerators	1.11	1.05	0.23	2.13
Clarifier (2)	0.70	0.50	0.53	0.52
DAF Thickeners (2)	0.35	0.40	0.46	0.58
Filters	0.37	0.17	0.19	0.19
Flash Mix*	0.03	0.02	0.02	0.02
Flocculation, Flotation*	0.05	0.02	0.02	0.02
Total Installed Cost	5.59	4.95	2.18	8.66

*Assume installed cost = 2 (equipment cost)

Water Purification Associates states that "...costs...are minimum; they do not include engineering fees...land...site preparation ...most foundation work...control building...laboratory."

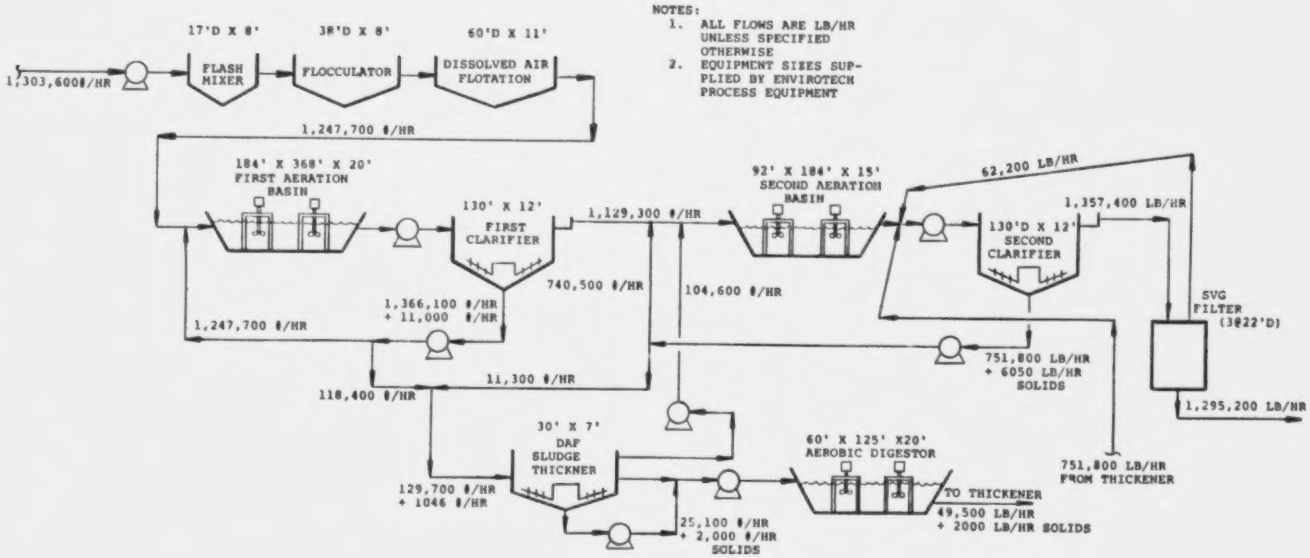
Also, the EPA/STP index for the above was 250, which corresponds to the year 1975. Costs should be increased for inflation alone on the order of 15 percent. The other omitted items would probably increase the costs at least to those obtained by using purchased equipment cost ÷ 0.1405 from the EPA/SCCT breakdown.

If we assume an overall relation, from the Engineering-Science, Inc. work mentioned earlier, of total cost = 3.12 (installed equipment cost), the total costs obtained would undoubtedly be high. Items not included in "installed equipment" in the Engineering-Sciences, Inc. method are mainly earthwork, concrete, control house, electrical equipment, connecting piping, and indirect costs. Indirect cost is 87 percent of installed cost, or 39 percent of base construction cost or 40.9 percent of the amount added to "installed equipment" costs. For aeration basin costs about half was stated as installed cost and half as earthwork and concrete. The factor of 3.12 could be adjusted downward for this. Removing half of the aeration basin cost in the four-case tabulation and applying the 3.12 factor yields the following capital costs:

Case I	$12,792,000
Case II	$11,107,000
Case III	$ 5,663,000
Case IV	$18,907,000

These figures still appear to be excessive, although not as high as those in the Braun report.

Additional information was obtained from Envirotech, including their estimates of installed cost of supplied equipment and more details on individual equipment, in order to determine the remainder of the facilities to be installed (mainly concrete basins and foundations) and to ascertain which items were not supplied. A process flowsheet for Case I, Figure 8-45, was then prepared



17'D X 8'
FLASH MIXER
38'D X 8'
FLOCCULATOR
60'D X 11'
DISSOLVED AIR FLOTATION
1,303,600#/HR
1,247,700 #/HR
184' X 368' X 20'
FIRST AERATION BASIN
130' X 12'
FIRST CLARIFIER
1,129,300 #/HR
740,500 #/HR
1,366,100 #/HR
+ 11,000 #/HR
1,247,700 #/HR
11,300 #/HR
118,400 #/HR
30' X 7'
DAF SLUDGE THICKNER
129,700 #/HR
+ 1046 #/HR
25,100 #/HR
+ 2,000 #/HR
SOLIDS
104,600 #/HR
92' X 184' X 15'
SECOND AERATION BASIN
62,200 LB/HR
130'D X 12'
SECOND CLARIFIER
1,357,400 LB/HR
SVG
FILTER
(3@22'D)
1,295,200 LB/HR
751,800 LB/HR
+ 6050 LB/HR
SOLIDS
751,800 LB/HR
FROM THICKENER
60' X 125' X20'
AEROBIC DIGESTOR
TO THICKENER
49,500 LB/HR
+ 2000 LB/HR SOLIDS

NOTES:
1. ALL FLOWS ARE LB/HR UNLESS SPECIFIED OTHERWISE
2. EQUIPMENT SIZES SUPPLIED BY ENVIROTECH PROCESS EQUIPMENT

Figure 8-45.- Flowsheet for Case I (Lurgi gasification).

for use by the Pullman Kellogg Cost Services Division in estimating the installed cost of the activated sludge treatment system. Details of the estimate are shown in TABLE 8-48.

Qualifications given by the Pullman Kellogg estimator were as follows:

- Site is assumed clear and level with no rock.
- Foundation quantities were estimated in-house using the equipment sketches furnished by Envirotech.
- Power is assumed to be available at battery limits at required voltage. No major substations or switch gear are included.
- Bulk commodities were ratioed from major equipment using best historical data.
- Contractor's costs (indirect material, engineering, insurance and overhead and profit) were based on normal general contractor practices and percentages.
- No forward escalation is included.
- No estimate accuracy is stated due to the preliminary nature of the data and the limited time spent. Estimate is for budget purposes only.

The contingency in the estimate is an allowance for undefined items. The contingency includes freight costs that are not otherwise estimated, although the estimator feels that freight could be considered to be included with the ratioing factors. The contingency allows for unforeseen costs such as additional engineering requirements and additional equipment.

Startup costs are not included in the estimate. These and other items such as interest during construction, working capital for initial operation and cost of land would increase the actual capital commitment required.

TABLE 8-48. ESTIMATED INVESTMENT FOR ACTIVATED SLUDGE TREATMENT SYSTEM

	Direct Material	Subcontracts*
Major Equipment	$ 1,025,000	
Site Preparation		$ 500,000
Steel Platforms, Walkways	100,000	
Piping	500,000	
Electrical	300,000	
Instruments	200,000	
Misc. Sumps, Lift Stations		50,000
Painting		100,000
Assembly, Earthwork, Foundations		1,945,000
Total Direct Material	$ 2,125,000	
Total Subcontracts	$ 2,595,000	
Construction Labor	$ 2,308,000	
Indirect Material	235,000	
Engineering	750 000	
Sales and Use Tax	150,000	
Insurance	50,000	
Total Bare Cost	$ 8,213,000	
Contractor Overhead and Profit	1,250,000	
Contingency	1,250,000	
Total Investment	$10,713,000	

*Includes some assembly costs furnished by Envirotech

The total investment cost shown for Case I may be rounded off to $10,000,000. The ratio of purchased equipment cost (furnished by Envirotech) to total cost is 0.112, a fair check with the average ratio in EPA's SCCT index of 0.1405 in a range of 0.113 to 0.203.

Using the equipment costs divided by 0.112 for Cases II, III and IV, their investments are:

Case II	$ 6,820,000
Case III	$ 4,250,000
Case IV	$11,500,000

Freight Costs

Since relative locations of equipment manufacturers' facilities and proposed coal conversion plants are unknown and variable, some orientation on freight costs is provided by the following from Pullman Kellogg's Expediting and Traffic Department.

Freight costs are a function of origin, destination, weight, volume and commodity. Equipment vendors generally quote costs of equipment "knocked down" so that it can be shipped to the construction site by truck or rail within the U.S. Freight costs are paid by the buyer in addition to quoted price.

As an example, shipping by truck from Los Angeles to Houston (about 1,800 miles would cost decreasing amounts per 100 pounds as the load neared a full truck load. A full truck load would be about 60,000 to 70,000 pounds as limited by highway regulations. Heavier haulers are available, but might be restricted in some states.

Weight Range	Cost per 100 lbs
1,000 - 2,000 lbs	$12.00
2,000 - 5,000 lbs	10.00
5,000 - 10,000 lbs	9.00
10,000 - Truck load	7.50
Full Truck Load	6.00

Thus, the freight cost on a 70,000 pound shipment would be $4,200 by truck from Los Angeles to Houston. Shipping pipe the same distance would cost about $1.00 per 100 pounds less.

Depending on dimensions of the shipment, it might be necessary to use rail freight beginning at 50,000 to 60,000 pounds. Rail rates are generally about $2.00 per 100 pounds higher than the truck rates cited. Therefore rail shipment of the 70,000 pounds of freight from Los Angeles to Houston would cost $5,600.

Ocean or barge freight would probably not prove practical unless the points of origin and destination were conveniently located along the same coast or navigable stream. For coal conversion plants these circumstances appear to be unlikely.

Costs for Side-Stream Softening of Cooling Tower Blowdown

An analysis of side-stream softening of cooling tower blowdown an its effect on a refinery cooling tower system is contained in "Economic Attractiveness of Side-Stream Softening," a paper by Marc Curtis, Calvin Morgan Associates, Houston that was presented at the Third Annual Conference in Treatment and Disposal of Industrial Wastewaters and Residues, April 18-20, 1978.

Equipment cost (presumably installed) for a 600 GPM (300,000 lb/hr) side-stream softening plant operating on cooling tower blowdown was presented as:

	Cost (1977)
Chemical Storage, handling and controls	$ 50,000
Softening reactor	100,000
Sand filter	20,000
Filter press	150,000
Pump station (3-600 GPM pumps, foundation, controls and fittings)	20,000
Contractor fees	350,000
Engineering	93,000
Piping ($3.00/in/ft, installed)	230,000
Total	$1,023,000

This installation would change the costs of operating the cooling tower as follows:

	Annual Costs (1977)	
Item	Before	After
Chromate	$ 273,000	$ 55,000
Zinc	37,000	23,000
Polyphosphate	49,000	0
Chlorine	48,000	24,000
$Ca(OH)_2$	0	77,000
Na_2CO_3	0	178,000
H_2SO_4	0	11,000
Raw Water costs	700,000	595,000
Water Treatment costs	700,000	595,000
Wastewater costs (1)	131,000	0
Pumping	6,000	3,000
Sludge disposal costs (2)	0	37,000
Labor	0	130,000
Maintenance (3)	0	19,000
Total Annual Cost	$1,944,000	$1,747,000

See notes on following page

(1) \$0.25/1000 gal based on activated sludge with 20 mg/l of Amoco PX-21 powdered activated carbon.

(2) Estimated at \$30 per ton.

(3) Estimated at 2% of equipment cost.

API Separator Cost

An equation for estimating the cost of API oil-water separators is given in the Stanford Research Institute report number 80 (804) as:

$$\text{Cost (1969)} = \$58{,}000\ (\text{MGD})^{0.84}$$

To this should be added 35 to 40 percent for "associated facilities and an escalation factor of 2.23 to yield:

$$\text{Total Installed Cost (1977)} = \$181{,}000(\text{MGD})^{0.84}$$

The report states that operating costs are 1.2¢ per 1,000 gallons. Applying the escalation factor increases the cost to the current level of 2.7¢ per 1,000 gallons. For a specific installation the operating cost must be adjusted for the percent of design throughput at which it operates.

Cost of a CPI separator is stated in the report to be only about 80 percent of the API separator cost.

Costs for Regenerating Granular Activated Carbon

A recent article from "Chemical Engineering" of January 2, 1978 entitled "Selecting a Thermal Regeneration System for Activated Carbon" by R.H. Zanitech and R.T. Lynch of Calgon Corp. presents current capital and operating costs for furnace reactivation systems.

Typical installed costs vs. capacity for multiple hearth furnaces and rotary kilns are:

Capacity, lb/day	Purchased Equipment	Installed Cost
5,000	$ 240,000	$ 850,000
10,000	360,000	1,270,000
30,000	770,000	2,700,000
60,000	1,200,000	4,200,000

Installation costs include foundations, structural equipment setting, electrical, instrumentation, site preparation, engineering, contractor overhead and profit and indirect costs. Utility and offsite facilities are assumed available at battery limits. The authors of the paper further estimated time required to design, procure, install, and start up the furnace to be two years if 12-month delivery time for equipment can be obtained.

Operating costs shown for reactivating spent granular carbon were shown as:

	Annual Operating Costs ($1,000)			
Capacity, lb carbon/day	5,000	10,000	30,000	60,000
Fuel: 8,000 Btu/hr at $3.00/MM Btu	45	90	265	525
Power at 3¢/KWH	10	20	50	80
Steam: 1.0 lb/lb at $4.00/Mlbs	10	15	45	90
Labor and Supervision	110	110	110	110
Makeup-Carbon: 7% at 57¢/lb	70	145	435	875
Maintenance: 8% of capital	70	100	215	335
General plant overhead	30	50	110	200
Total Operating Cost	345	530	1,230	2,215
Operating Cost, ¢/lb carbon regenerated	18.8	14.5	12.2	11.0

It is evident that the main costs aside from amortization are fuel, makeup carbon and maintenance.

Dry Discharge Method - Cost Data

Dry solids discharge was required at the Peoples Gas SNG plant at Elwood, Ill. due to restricted disposal alternatives. This plant was constructed by Pullman Kellogg. The water treating system was designed by a water technology consultant and Illinois Water Treatment Company. Reference is "Operating Experiences with a Zero Discharge Deionizer," by Krol, Jones, Picht, and Martin (659). The paper was presented at the 38th Annual Meeting of the International Water Conference, Engineers Society of Western Pennsylvania, and printed by Illinois Water Treatment Company in 1977.

Components of the system are weak acid cation treatment to produce cooling water makeup, followed by strong acid cation and weak base anion beds. The treated water is then joined by waste evaporator condensate and decarbonated by aeration. The deaerated stream is combined with process condensate and the combined flow is polished by two-bed strong acid, strong base ion exchangers. Regeneration wastes are separately evaporated in two 3-stage evaporators to about 20 percent solids and then spray-dried, using naphtha for fuel. A bag filter follows the spray drier to collect the dry solids. HCl is used in regeneration instead of H_2SO_4 to avoid calcium sulfate scale formation and consequent heat exchanger fouling in the evaporators.

Operating costs of the first set of ion exchangers, that produce cooling tower makeup, were (1977) $1.05/.1,000 gallons. Operating costs of the total system, producing boiler feedwater quality water, were (1977) $2.69/ 1,000 gallons, consisting of (per 1,000 gallons) energy charges at 45.3¢, chemicals at 72.5¢, labor and maintenance at 48.4¢, salt disposal at 4.1¢ and capital and insurance charges at 98.7¢.

The basis for these operating costs was:

Electricity	2.17¢/KWH
Synthetic Gas	0.4¢/CF
Naphtha	32¢/gal
HCl (30%)	2.8¢/lb
NaOH (100%)	9.26¢/lb
W.A. Cation Resin	$86.00/CF
S.A. Cation Resin	$40.00/CF
W.B. Anion Resin	$102.00/CF
S.A. Cation Resin (10%)	$45.10/CF
S.B. Anion Resin	$112.00/CF
Labor and Maintenance	$12.00/hr
Capital Charges	$10%/year

The labor and maintenance figure was projected. In actual operation the spray dryer needed cleaning every three days, more frequently than anticipated, and increased the labor and maintenance charges. The condition is considered to be correctable.

An interesting note: in operation it was found to be unnecessary to blow down the cooling tower because of the efficiency of the weak acid cation treatment. Provision had been made in the plant design for blow down to be fed to one of the evaporators and elimination of this stream reduced the evaporator load.

Operating Costs of Water Treating Processes

The open literature and government reports contain many references quoting operating costs for water treating processes. Most of these costs are presented as cents/thousand gallons wastewater treated. Generally these costs purport to include amortization. As will be illustrated in this section, use of such data is risky since many are out of date, the bases for utility costs and amortization are hazy at best, and they mask the wide variation that can be encountered due to capacity, feed, and effluent variations.

Complete Operating Cost Estimates--

The only truly unassailable method of obtaining and portraying operating costs is to proceed on a single case basis where feed capacity and composition are known, all utility requirements are stated, values of utilities are stated and all capital-related items are clearly identified. Such a procedure was possible on several processes where licensors supplied data based on feed and effluent compositions specified by Pullman Kellogg. The cost analyses for Phenosolvan, Chevron Stripping and Ammonia Recovery and Zimpro Wet Air Oxidation are shown in Table 8-49.

Utility costs are based on a Pullman Kellogg article in Chemical Engineering, January 21, 1974, "Energy Conservation in New Plant Design," by J. B. Fleming, J. R. Lambrix, and M. R. Smith. Utility prices quoted in this article were projected for the period 1975-1980:

Fuel	Natural Gas	\$0.80 /MM Btu
	Light Hydrocarbons	\$1.50 /MM Btu
	No. 1 and No. 2 Fuel Oil	\$1.30 /MM Btu
	Bunker C Fuel Oil	\$1.20 /MM Btu
Steam	500 to 750 psig	\$2.00 /1,000 lbs
	20 to 200 psig	\$0.50 /1,000 lbs
Electricity		\$0.015/KWH
Cooling Water		\$0.04 /1,000 gals

TABLE 8-49. OPERATING COSTS OF SELECTED WASTEWATER TREATMENT PROCESSES

	Phenosolvan	Chevron Stripping and Ammonia Recovery	Zimpro Wet Air Oxidation of Liquefaction Water
Capacity, MGD	3.6	3.6	0.533
Capital Investment, $MM	13.871	12.00	14.00
Steam, 1,000 lbs/day	288 to 972	8,760	(Produce 720)
Cooling Water, 1,000 gals/day	5,400	5,472	2,160
Electric Power, KWH/day	14,400 to 21,600	43,200	138,700
Labor, man-shifts/day	N.A.	0.5	3
Isopropyl Ether, lbs/1,000 gal (1)	0.6 to 1.2	-	-
Caustic Soda (20% soln.), lbs/hr (2)	-	2,500	-
Corrosion Inhibitor, cost per day	-	$1,000	-
Cost of Utilities, Labor, Chemicals,			
per day	$1,058 to 2,387	$9,050	$2,407
per 1,000 gals	$0.29 to 0.67	$2.51	$4.51
Credit for Steam Produced,			
per day	-	-	($1,440)
Per 1,000 gals	-	-	($2.70)
Capital-Related Items (3),			
per day	$5,700	$4,932	$5,753
per 1,000 gals	$1.58	$1.37	$10.79
Total Treatment Cost,			
per day	$6,758 to 8,087	$13,982	$6,720
per 1,000 gals	$1.87 to 2.25	$3.88	$12.60

---

(1) At $0.19 per pound
(2) At $0.23 per pound 100 percent NaOH
(3) Includes amortization, maintenance, insurance, etc.

To this list should be added the cost of boiler feed water, which we assumed at $0.10/1,000 gallons.

Although some of the utility costs may be open to argument in various sections of the U.S., they are being used in this analysis for the sake of consistency. Local values can be substituted, provided that the utility quantity is stated.

Operating labor rate was taken as $80 per man per shift.

Chemical prices were taken from the "Chemical Marketing Reporter."

American Lurgi and Chevron furnished data from which the treatment costs of smaller capacity plants were calculated:

	Phenosolvan	Chevron Stripping and Ammonia Recovery
Capacity, MGD	0.533	1.08
Investment, $MM	5.07	6.00

	Phenosolvan per day	Phenosolvan per 1,000 gals	Chevron per day	Chevron per 1,000 gals
Operating Costs	$ 157-353	$0.29-0.67	$3,759	$3.51
Capital Costs	2,083	3.91	$2,466	$2.28
Total Treatment Cost	$2,240-2,436	$4.20-4.58	$6,225	$5.79

Comparison of the total treatments costs in $ per 1000 gallons in the larger plants and the smaller plants indicates the variation in costs with capacity and emphasizes the danger of using the undefined "typical" costs that are often quoted in the literature.

Adequate data were supplied for operating cost calculations for biological oxidation using powdered activated carbon and regeneration of the carbon with Zimpro's wet oxidation technique. Zimpro supplied the basic figures and has cooperated with DuPont on the overall process. The process will be referred to by the DuPont simplified designation PACT (Powdered Activated Carbon Treatment). Costs are shown in Table 8-50. Note that phenol was extracted from the larger flow prior to biological oxidation.

The striking aspect in comparison of the two systems is that despite the difference in capacity, each plant processes about the same total amount of BOD per day, the total treatment costs per day are virtually the same and the treatment costs per 1,000 gallons are in about the same ratio as are the BOD's in the feeds. From this comparison the conclusion may be drawn that for the PACT process with carbon regeneration the total treatment costs are directly proportional to the BOD content of the feed and very much less dependent on the volumetric throughput.

Partial Operating Cost Estimates--

For other water treatment processes only partial operating costs could be developed from information received. Comments on these costs follow.

Flotation--The electric power required to operate the air compressor and pumps can be calculated:

Capacity, MGD		4.03	0.85	0.53
Electric Power,	per 1,000 gals	$0.004	$0.006	$0.005
	per day	$17.30	$5.50	$2.90

The required alum and polyelectrolyte quantities would have to be determined experimentally. If pH adjustment were necessary, the cost of the adjustment chemicals would be chargeable to the

TABLE 8-50. OPERATING COSTS FOR BIOLOGICAL OXIDATION WITH POWDERED ACTIVATED CARBON

	Capacity 3.6 MGD		Capacity 1.086 MGD	
Feed Source	Lurgi Wastewater		SRC-II Combined Wastewater	
Prior Phenol Extraction?	Yes		No	
BOD, mg/l	1,700		5,800	
Electric Power, KWH/day	59,000		54,000	
Labor, man-shifts/day	6		6	
Activated Carbon, $0.28/lb, lbs/day	3,000		3,000	
Polymer, $2.50/lb, lbs/day	68		18	
Phosphorus, $0.235/lb, lbs/day	510		526	
	per day	per 1,000 gals	per day	per 1,000 gals
Cost of Utilities,	$2,495	$0.69	$2,298	$2.11
Capital-Related Items*	$4,110	1.14	$3,904	3.59
Total Treatment Cost	$6,605	$1.83	$6,202	$5.70

*At 15 percent of capital per year

process. As an example of these costs, we could assume that 100 mg/l of alum is used, as at the Fort Lewis, Wash. 50 TPD SRC plant. This cost works out to be $205.50 per day or $0.051 per 1000 gallons for the 4.03 MGD plant. This is more than ten times the cost of electricity and it is obvious that chemical costs will be controlling in the flotation process.

Biological Oxidation--Electric power to operate the aerator motors must be considered. Envirotech furnished the data for the four cases described earlier in the report. Power requirement is a direct function of BOD removed:

	Lurgi		Combined Liquefaction Wastewater	
Capacity, MGD	4.03	0.85	0.80	0.53
BOD removed, lbs/hr				
1st stage	1,720	480	4,230	1,880
2nd stage	181	51	212	94
Total	1,901	531	4,442	1,974
Electric Power Costs				
Per day	$358	$84	$676	$358
Per 1,000 gals	$ 0.094	$ 0.096	$ 0.85	$ 0.678
Per 1,000 lb BOD removed	$ 20.20	$17.31	$ 18.57	$ 20.88

Other costs of biological oxidation are better illustrated in the examples furnished by Zimpro for the PACT process and include air compressor power, phosphorus costs, labor and amortization.

Evaporation--Envirotech (Goslin) furnished the basis for development of the following operating costs:

Feed Rate, lbs/hr	283,700		315,580	
	per day	per 1,000 gals	per day	per 1,000 gals
Steam, 18 to 150 psig,$0.50/ 1,000 lbs	$732	$0.90	$814	$0.90
Electric Power, $0.015/KWH	24	0.03	30	0.03
Cooling Water, $0.04/ 1,000 gal	122	0.15	137	0.15
Total Operating Cost	$878	$1.08	$981	$1.08

This is one process where capacity governs operating cost directly, since chemical reactions or additions are involved only when it is necessary to control scale formation.

Demineralization--For demineralization (ion exchange) of water having the analysis used by Braun for the western coal gasification designs, L*A Water furnished the basis for these costs:

Capacity	1,137 GPM (1.64 MGD)	
Regenerations per day	6.82	
Sulfuric Acid	$209/day	($0.13/1,000 gals)
Caustic	$1605/day	($0.98/1,000 gals)
Electricity	$3.07/day	(0.003/1,000 gals)

Again it is obvious that chemical costs are governing in the operating costs. Chemical costs vary directly with the necessity for regeneration of the resin beds, which in turn is a function of the feed water composition (amounts of anions and cations to be removed) and the size of the resin beds.

Reverse Osmosis--The operating cost is governed largely by pumping cost. L*A Water furnished the basis for the following:

Effluent Flow	1,540 gpm (2.22 MGD)
Electric Power, \$0.015/KWH, per day	\$244
per 1,000 gals	\$0.11

Greater costs than the above could be incurred by the necessity for filtration preceding the reverse osmosis unit. No costs were supplied for operating the filter unit.

Demineralization and Reverse Osmosis--Probably the best operating costs to use are those appearing in a Dow Chemical advertisement (Chemical Engineering, April 10, 1978). Total operating costs, including capital factors, are shown as follows:

Feed Rate	500,000 gals per day		
Total Dissolved Solids (ppm as $CaCO_3$)	200	360	600
Ion Exchange Cost, per 1,000 gals	\$1.15	\$1.60	\$2.50
Reverse Osmosis cost, per 1,000 gals	\$0.32	\$0.32	\$0.34
Reverse Osmosis Preceding Ion Exchange Costs, per 1,000 gals	\$1.60	\$1.60	\$1.65

Budget Cost Estimates Received from Licensors and Vendors

Stripping and Ammonia Recovery in the Chevron WWT Process--
Cost data for stripping of H_2S, CO_2, and NH_3 and ammonia recovery were furnished to us by the Chevron Research Company for their licensed WWT process, based on flows and compositions for three cases: (1) Lurgi (p/o/t-producing gasification), (2) Bi-Gas/Kopper-Totzek (non-p/o/t producing gasification) and (3) SRC-II (liquefaction plus non-p/o/t-producing gasification to produce fuel gas and hydrogen). In Cases 1 and 3, phenol extraction was simulated (by calculation) preceding their stripping process. Compositions and flows are discussed in a following section of this report entitled "Integrated Schemes for Wastewater Treatment."

Figure 8-46 is a block flow diagram of the Chevron Research process. The process description submitted was as follows:

> The Chevron Waste Water Treating Process (WWT Process) is a patented process for treating foul water streams from petroleum refineries, coal processing plants, and synthetic fuel plants to recover and separate high purity ammonia and hydrogen sulfide and to recover clean water for reuse or discharge. It is particularly applicable to foul water streams from hydrotreaters and hydrocrackers. Ammonia recovery from foul water effluents from FCC and coking plants also can be considered.
>
> The WWT process consists basically of two distillation columns. Overhead product of one column is H_2S gas containing less than 30 ppm by weight of ammonia. The overhead product from the other distillation column is ammonia containing less than 5 ppm by weight of H_2S. This ammonia can be liquefied and sold as anhydrous material or sold in aqueous form. It meets all agricultural grade specifications. The stripped water contains less than 50 ppm of NH_3 and 5 ppm H_2S by weight. It can be

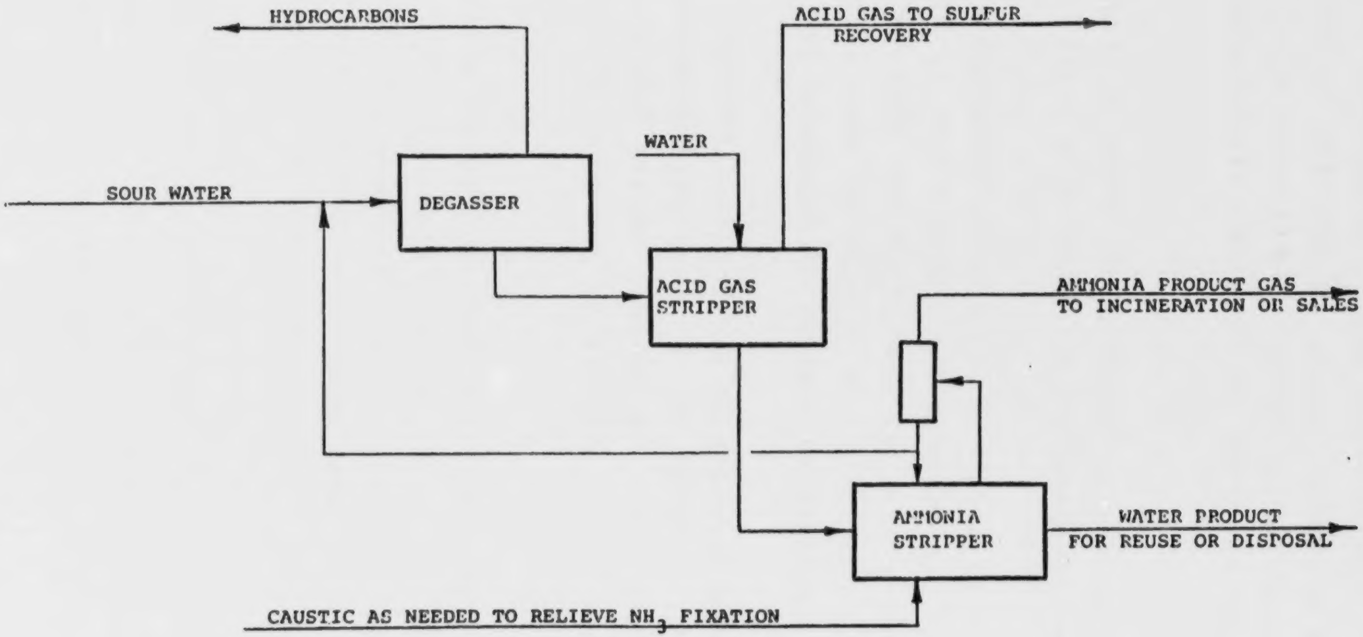

HYDROCARBONS
ACID GAS TO SULFUR RECOVERY
WATER
SOUR WATER
DEGASSER
ACID GAS STRIPPER
AMMONIA PRODUCT GAS TO INCINERATION OR SALES
AMMONIA STRIPPER
WATER PRODUCT FOR REUSE OR DISPOSAL
CAUSTIC AS NEEDED TO RELIEVE NH_3 FIXATION

Figure 8-46. The Chevron WWT process.

recycled to refinery process units, except for a small bleed stream which is normally taken to prevent buildup of trace inpurities. Reduction in refinery effluents and clean make-up water demand have made the WWT Process attractive in several situations.

Thirteen WWT plants are in operation or under construction, one of which has been in operation for about nine years.

TABLES 8-51, 8-52, 8-53 and 8-54 illustrate investment and operating costs furnished by Chevron for Cases 1 and 3. It was stated that ammonia recovery was not justified for Case 2; however, it will be necessary to install the stripping towers for this Case, even if ammonia is incinerated instead of recovered for sale.

Stripping and Ammonia Recovery in the Phosam-W Process--
An alternate for the Chevron WWT Process is the Phosam-W process, licensed by U.S. Steel Co. Stripping, separation of CO_2 and H_2S from NH_3, and recovery of ammonia by ammonium phosphate solvent are included.

Costs for Phosam-W systems have been quoted by Water Purificaton Associates (480) and by Ralph M. Parsons (814). C.F. Braun (294,295,296) also concluded that Phosam-W was the process of choice. Cost data from these references follow. The Braun cost is pro-rated down from 7.3 MGD to 3.6 MGD using a 0.6 exponent.

	W.P.A.		Parsons	Braun
Capacity, MGD	3.6	1.06	1.064	3.6
Major Equipment Costs, $MM	7.2	2.75	1.589	5.5
Total Investment Cost, $MM	Not Stated		5.553	24
Royalty, $MM	1.0	-		
Operating Cost, including amortization, $/ 1,000 gal	4.02	4.25	-	-

TABLE 8-51. INVESTMENT AND UTILITY ESTIMATES FOR THE CHEVRON WWT PROCESS, PULLMAN KELLOGG CASE 1

Feed	
Nominal Sour Water Rate, gpm	2,500
NH_3, Wt %	1.2
H_2S, Wt %	0.051
CO_2, Wt %	1.68
Products (1)	
Acid Gas ($H_2S + CO_2$), Short Ton/Day	260
Ammonia, Short Ton/Day	180
Water, gpm (60° F)	2,430
Investment (2)	
Total Installed Cost	$12,000,000
Utilities (3)	
Steam (With Returnable Condensate)	
150 psig, Lb/Hr	195,000
50 psig, Lb/Hr	170,000
Electrical Power, kw	1,800
Cooling Water, gpm	3,800
Maintenance, Cost/Year	$360,000
Operating Labor, Man/Shift	1/2
Onplot Area, Sq. ft.	14,000
Chemicals (4)	
Corrosion Inhibitor, Cost/Year	$370,000
Caustic (20%), Lb/Hr	2,500

(1) Product compositions and conditions are given in TABLE 8-52
(2) Investment is based on current prices for a U.S. West Coast site. Investment figure does not include a required feed tank, product handling facilities, or other offplot requirements
(3) Utility consumption is based on maximized use of air cooling. Maintenance is based on 3% of investment
(4) Caustic estimated from the fatty acid content of the feed

TABLE 8-52. PRODUCT COMPOSITIONS AND CONDITIONS FOR THE CHEVRON WWT PROCESS, PULLMAN KELLOGG CASE 1

Acid Gas Product

The acid gas product (CO_2-H_2S mixture) will consist mainly of carbon dioxide and will have a maximum ammonia content of 250 ppm by weight (or 50 ppm at the expense of 70 gpm deaerated condensate consumption). The acid gas will be saturated with water at 100°F and the process pressure. Conditions at plot limit are a maximum temperature of 120°F and a pressure ranging from 10 psig up to 120 psig as required.

Ammonia Product

The ammonia product is produced in the anhydrous liquid form. It will have a maximum hydrogen sulfide plus carbon dioxide content of 5 ppm by weight and a water content of 0.05 percent by weight. Conditions at plot limit are a maximum temperature of 100°F and a minimum pressure of 200 psig.

Water Product

The water product is essentially pure water containing traces of phenols and salts which entered with the feed. The ammonia content will be a maximum of 50 ppm by weight, and the free acid gas content will be a maximum of 10 ppm by weight. Plot limit temperature is 140°F.

TABLE 8-53. INVESTMENT AND UTILITY ESTIMATES FOR THE CHEVRON WWT PROCESS, PULLMAN KELLOGG CASE 3

Feed	
Nominal Sour Water Rate, gpm	750
NH_3, Wt %	2
H_2S, Wt %	1.93
CO , Wt %	2.49
Products (1)	
Acid Gas (H_2S + CO_2), Short Ton/Day	200
Ammonia, Short Ton/Day	90
Water, gpm (60° F)	700
Investment (2)	
Total Installed Cost	$6,000,000
Utilities (3)	
Steam (With Returnable Condensate) 150 psig, Lb/Hr	160,000
Electrical Power, kw	1,100
Cooling Water, gpm	3,500
Maintenance, Cost/Year	$180,000
Operating Labor, Man/Shift	1/2
Onplot Area, Sq. ft.	10,000
Chemicals (4)	
Corrosion Inhibitor, Cost/Year	$140,000
Caustic (20%), Lb/Hr	740

(1) Product compositions and conditions are given in TABLE 8-54

(2) Investment is based on current prices for a U.S. West Coast site. Investment figure does not include a required feed tank, product handling facilities, or other offplot requirements

(3) Utility consumption is based on maximized use of air cooling. Maintenance is based on 3% of investment

(4) Caustic estimated from the fatty acid content of the feed

TABLE 8-54. PRODUCT COMPOSITIONS AND CONDITIONS FOR THE CHEVRON WWT PROCESS, PULLMAN KELLOGG CASE 3

Acid Gas Product

The acid gas product (CO_2-H_2S mixture) will have a maximum ammonia content of 250 ppm by weight (or 50 ppm at the expense of 60 gpm deaerated condensate consumption). The acid gas will be saturated with water at 100°F and the process pressure. Conditions at plot limit are a maximum temperature of 120°F and a pressure ranging from 10 psig up to 120 psig as required.

Ammonia Product

The ammonia product is produced in the anhydrous liquid form. It will have a maximum hydrogen sulfide plus carbon dioxide content of 5 ppm by weight and a water content of 0.05 percent by weight. Conditions at plot limit are a maximum temperature of 100°F and a minimum pressure of 200 psig.

Water Product

The water product is essentially pure water containing traces of phenols and salts which entered with the feed. The ammonia content will be a maximum of 50 ppm by weight, and the free acid gas content will be a maximum of 10 ppm by weight. Plot limit temperature is 140°F.

These data are not directly comparable, since bases are not clearly understood for the W.P.A. costs. Although described on pages 319 of Reference 480 as "capital cost," on page 340 the numbers in the 3.6 MGD column are stated as "total equipment." W.P.A. costs are for 1975 (multiply by 1.12, minimum, for escalation). Parsons costs are clearly stated as fourth quarter 1975 (multiply by 1.04 to 1.112, where 1.04 is the Pullman Kellogg process plants escalation factor and 1.112 is the EPA/SCCT municipal sewage plant correction).

The Braun references (294, 295, 296) give sour water stripping and ammonia recovery costs for March 1, 1976 for Bi-Gas and CO_2 Acceptor:

	Bi-Gas	CO_2 Acceptor
Capacity, MGD	7.14	0.98
Total Engineered Equipment, \$MM	8.4	2.65
Installed Cost, \$MM	37.0	18.00

Above included a separate stripper plus a package cost for Phosam-W ammonia recovery (which presumably includes paid-up royalty). Ammonia storage spheres are also included. These costs are nevertheless difficult to reconcile with the Parsons or W.P.A. costs in the previous tabulation.

Pullman Kellogg requested costs directly from USS Engineers and Consultants for the three cases described in the Chevron WWT process cost analysis. Capacity for Case 1 and Case 2 was 3.6 MGD. Capacity for Case 3 was 1.09 MGD. Compositions were the same as those sent to Chevron.

USS Engineers and Consultants, Inc. replied as follows:

> Our Phosam-W process engineers have reviewed the three sour

wastewater cases presented in your letter of April 21, 1978, and have concluded that the Phosam-W process is applicable to these feed streams. However, to date, we have not designed a Phosam-W plant which is preceded by a lime softening and stabilization step. We assume that the feed water stream will be free of fouling agents which may cause a problem in the sour water stripper.

Our specific comments and cost estimates for each of the three cases presented are as follows:

Case 1 can be handled by a typical Phosam-W plant, which consists of a sour water stripper followed by absorption of ammonia from the stripped vapor by an ammonium phosphate solution. The phosphate solution is then stripped to produce aqueous ammonia for fractionation to anhydrous ammonia product for sale. The cost estimate for an installed battery-limits plant, without ammonia storage, to produce 180 tons per day (TPD) of anhydrous ammonia is approximately $14 million.

Case 2 is uneconomical because of the extremely low feed concentration (530 ppm) and the low potential production rate (8 TPD of ammonia). However, if ammonia removal is required for other than economic reasons (i.e., environmental standards), a Phosam-W plant could be designed for this case. For your information, it is our understanding that other proposed coal gasification plants would use waters of this nature as cooling tower makeup.

Case 3 is an attractive case for a Phosam-W plant. The cost estimate for an installed battery-limits plant, without ammonia storage, to produce 90 TPD of anhydrous ammonia is approximately $9 million.

A nonconfidential design document can be prepared for Case 1 or Case 3, which are the most attractive feeds. This work would take approximately five to eight weeks to complete and would cost $10,000.

Capital costs of Chevron WWT and Phosam-W plants are compared on the same basis as quoted to Pullman Kellogg:

	Chevron WWT	Phosam-W
Case 1 (Lurgi, 3.6 MGD)	$12,000,000	$14,000,000
Case 3 (SRC, 1.086 MGD)	6,000,000	9,000,000

If the relationship

$$\text{Capital}_1\ (\text{Capacity}_2/\text{Capacity}_1)^x = \text{Capital}_2$$

can be assumed to hold, on the basis of two points of information, it is of interest to note that $x = 0.58$ for the Chevron plants and only 0.37 for the Phosam-W plants. This seems to indicate that the Phosam-W process is more capital-intensive than the Chevron process, with the result that small Phosam-W plants are penalized. As capacity increases, however, the difference between the capital costs of the two designs decreases until, at 7.4 MGD, the capital costs are equal. Above 7.4 MGD, Phosam-W apparently has the advantage.

Biological Oxidation with Powdered Carbon/Wet Air Oxidation--Zimpro, Inc. supplied us with estimated investment and operating costs for processing four streams.

Stream 1--This is the raw liquefaction dissolver wastewater containing high concentrations of phenols and other organics (BOD = 52,700, COD = 88,600). No prior processing was contemplated in order to ascertain whether wet air oxidation could be substituted for conventional processing. Zimpro's comments for this stream follow. Zimpro's estimates of capital and operating costs are shown in TABLE 8-55.

> The following costs are for a Wet Air Oxidation system consisting of three parallel 125 gallon per minute units designed to handle the wastewater described in your letter. I might add that this estimated capital cost is for a complete system installed, less buildings and foundations. The operating costs and steam production (150 psig saturated) are the total for the three units.
>
> The oxidized waste will contain some residual organic material which will likely necessitate subsequent biological treatment. However, the residual organics contained in the oxidized waste will be readily biodegradable and non-toxic. From the estimated residual BOD and COD, the size of the biological system for handling the residual organics would be substantially reduced since most of the organic constituents were destroyed during Wet Air Oxidation. It is also possible to utilize this Wet Air Oxidation system for handling the waste sludge production from the biological treatment system used for post treating the oxidized waste.
>
> Estimated BOD in the discharge is 7,000 mg/l (87 percent reduction). Estimated COD in the discharge is 9,000 mg/l (90 percent reduction).

Figure 8-47 is a schematic illustration of the processing for Stream 1.

Stream 2--This is the total of wastewater from liquefaction combined with wastewater from the process and the fuel gas producers. It has been stripped and the oil has been

TABLE 8-55. CAPITAL AND OPERATING COSTS FOR BIOLOGICAL OXIDATION WITH POWDERED ACTIVATED CARBON AND WET AIR OXIDATION

Stream 1: Raw liquefaction dissolver wastewater: Wet Air Oxidation System
Stream 2: Combined liquefaction feed to biological oxidation: PACT plus Wet Air Oxidation of Sludge
Stream 3: Combined sour water from Lurgi gasification

	Stream 1	Stream 2	Stream 3
Capacity, MGD	0.54	1.086	4.03
Estimated Capital Investment, $MM	$12 to 16	$8 to 11	$8.5 to 11.5
	Cost per Day	Cost per Day	Cost per Day
Electric Power at $0.025/KWH	$4592	$1350	$1475
Maintenance Labor and Materials	600	400	450
Cooling Water at $0.05/1000 gals.	108	-	
Operating Labor at $80/man-shift	720	480	480
Polymer at $2.50/lb	-	45	170
Makeup Carbon at $0.28/lb	-	840	840
Total Operating Cost	$6020	$3115	$3415
Credit for Steam at $3.00/1000 lbs.	$2160		
Net Operating Cost	$3860		

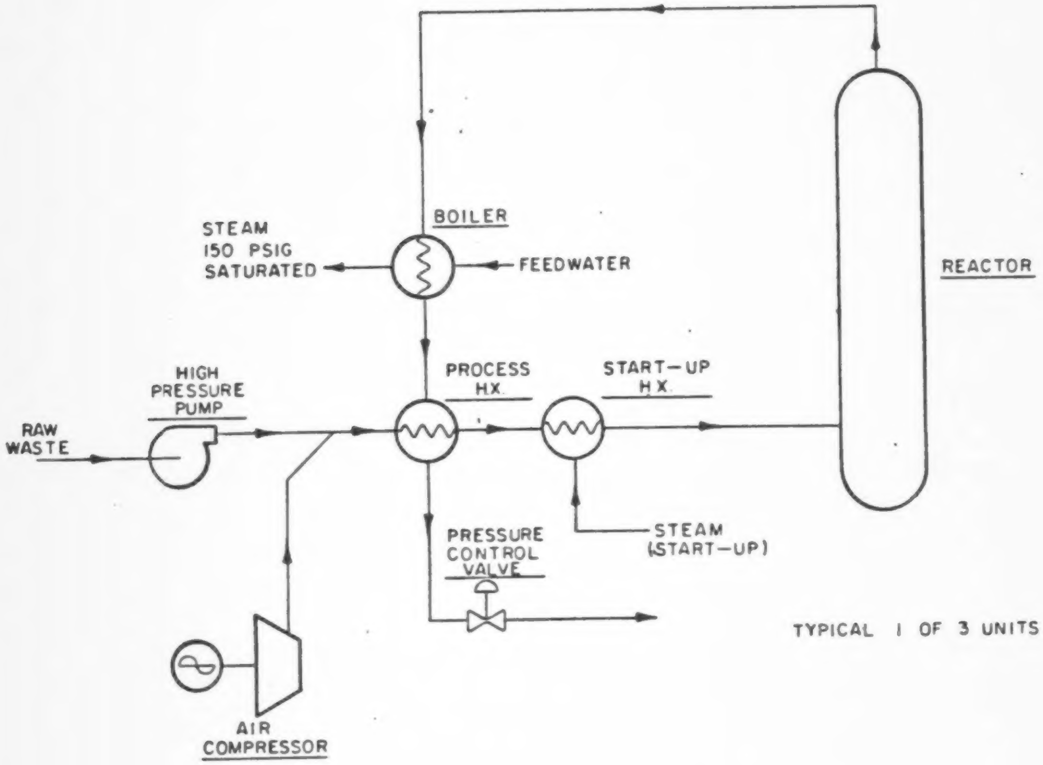

BOILER
STEAM
150 PSIG
SATURATED
FEEDWATER
REACTOR
HIGH
PRESSURE
PUMP
PROCESS
HX.
START-UP
HX.
RAW
WASTE
PRESSURE
CONTROL
VALVE
STEAM
(START-UP)
TYPICAL 1 OF 3 UNITS
AIR
COMPRESSOR

Figure 8-47. Schematic W.A.O. for stream # 1

separated as would be done for biological oxidation. Zimpro proposes a biological system employing powdered carbon with sludge oxidized by wet air oxidation for this stream.

For Streams 2 and 3, we are recommending a two-stage Wastewater Reclamation System (WRS). This system includes a Wet Air Regeneration system for regenerating the spent activated carbon and oxidizing the waste biological solids associated with the spent carbon.

Zimpro has done considerable work on coke oven flushing liquors which are quite similar in terms of constituent pollutants to the waste you describe. Based on this experience, we are confident that the WRS will produce a high quality effluent. We would expect the following reductions: COD >93 percent; BOD >99.5 percent; cyanides >99 percent; phenols virtually 100 percent; thiocyanates 90 percent.

In reviewing your list of components in the waste, I have concluded that more ammonia and phosphorus will be needed to meet the nutrient requirements of the waste. I can account for 61.5 mg/l of NH_3-N in the waste in the form of CN, SCN and NH_3. Generally, the ratio of nutrients to BOD is 1-5-100 phosphorus-nitrogen-BOD. Based on this, the nitrogen requirement would be 290 mg/l. This means an additional ammonia content of 228.5 mg/l would be required. Likewise, phosphorus in the amount of 58 mg/l would generally be considered necessary.

As you know, these nutrient requirements are not unique to the WRS since proper nutrient concentrations will be required for any biological treatment system.

Zimpro's estimates of capital and operating costs are shown in

TABLE 8-55.

Figure 8-48 is a schematic illustrating the system for both Stream 2 and Stream 3.

Stream 3--This is the combined sour water from Lurgi gasification after oil separation, stripping, and phenol extraction; i.e. the stream ready to undergo biological oxidation in the conventional manner. Zimpro's comments follow:

> The expected performance of the WRS for Stream 3 is the same as for Stream 2. Stream 3 is essentially the same in terms of lbs/day of COD and BOD as Stream 2. Therefore, the costs associated with a WRS system are nearly the same as Stream 2. One difference associated with the higher flow rate is the increase in clarifier size which is the major contributor to the slight increase in capital cost.
>
> Stream 3 seems to have an adequate nitrogen supply, but may be phosphorus deficient. Assuming no phosphorus is present in the waste, I would estimate that a phosphorus dose rate of 17 mg/l would be required.

Zimpro's estimates of capital and operating costs are shown in TABLE 8-55.

Streams 4 and 5--These were estimates of conventional biological oxidation sludge for liquefaction and Lurgi gasification, respectively. The intent was to determine whether wet air oxidation would be a viable alternate to sending this sludge to our central incinerator/boiler. Zimpro's comments follow:

> Streams 4 and 5 consist of waste sludge from the Liquefaction and Lurgi Gasification Biox systems. As we discussed, I

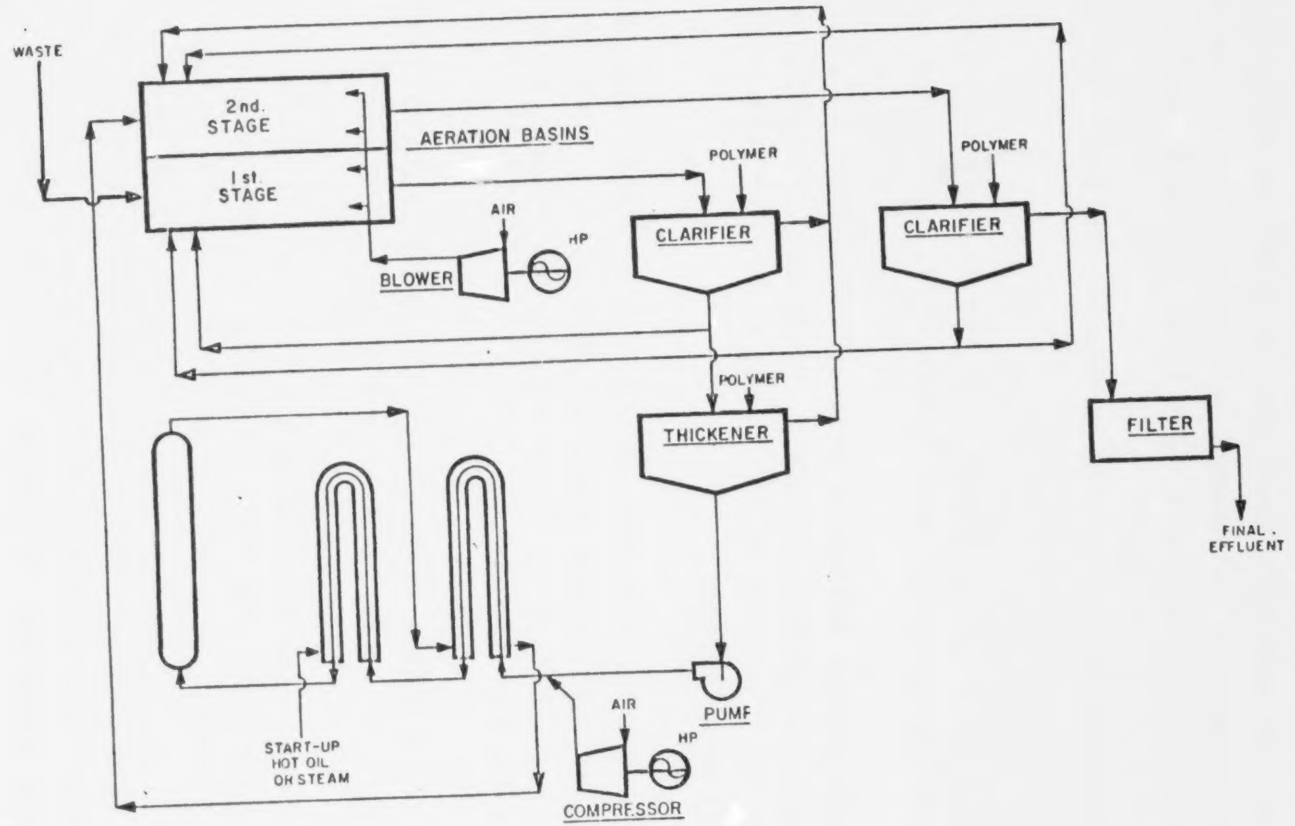

WASTE
2nd. STAGE
1 st. STAGE
AERATION BASINS
AIR
HP
BLOWER
POLYMER
CLARIFIER
POLYMER
CLARIFIER
POLYMER
THICKENER
FILTER
FINAL . EFFLUENT
START-UP HOT OIL OR STEAM
AIR
HP
COMPRESSOR
PUMF

Figure 8-48. Schematic for WRS streams #2 & #3.

believe your estimated sludge flow rates are a bit low, based on the influent BOD to the two Biox systems. It is true that long solids retention times will produce low sludge yields due to the effects of endogenous respiration, but I still believe the yields will be higher than those you have listed.

My estimate of the solids produced in each of the above systems is as follows:

Liquefaction Biox System - 15,750 lbs/day dry solids
Lurgi Gasification Biox System - 17,100 lbs/day dry solids

We would assume the above waste solids would be thickened to at least 3.5 percent solids by means of a flotation thickener prior to feeding the streams to the thermal sludge conditioning system.

For your reference, we do not supply flotation thickening equipment, but I have estimated the cost of such for systems capable of thickening the above streams from 0.94 percent to 3.5 percent as follows:

Liquefaction Biox or Lurgi Gasification Biox sludge thickener - capital cost installed: $180,000

As you know, the Process we will propose for this is a thermal sludge conditioning unit which is a Low Pressure Oxidation (LPO) system. With an LPO system, we will do a low degree of oxidation (about 5 percent in terms of COD reduction) and raise the temperature of the sludge to about 180°C. This combination of oxidation and heat will dramatically improve the dewatering and thickening characteristics.

Following the LPO system, we would propose a filter press to allow final dewatering of the solids to about a moisture content of less than 60 percent. The resultant filter cake can be easily disposed of on land with no potential public nuisance problems or, alternatively, it could be mixed with coal and burned in a coal fired boiler.

The capital cost and operating costs for an LPO system and filter press installed, not including buildings and foundations are shown in TABLE 8-56.

The schematic for processing of Streams 4 and 5 is shown in Figure 8-49.

Zimpro submitted a technical paper describing a "Wet Oxidation Boiler Incinerator" which could be considered as an alternate to our central incinerator/boiler.

UNOX Biological Oxidation: Quote from Union Carbide--
Through telephone contact and correspondence an equipment cost and design information were obtained from Union Carbide, who are licensors of the UNOX process which employs high purity oxygen instead of air in biological oxidation. The presentation in TABLE 8-57 was taken from their response. Case I refers to wastewater from a gasification process typified by Lurgi (producing p/o/t) after phenol extraction, stripping, coagulation, and flotation. Case III refers to wastewater from a liquefaction dissolver similarly treated. Quotation from Union Carbide's reply follows.

Effluent Standards--In both cases, the wastewater effluent is required to meet the standards shown in TABLE 8-57. The standards for TDS, COD, and phenol are unusually restrictive and may not be obtainable by secondary treatment. As a comparison, the Illinois EPA has set a TDS level of 1,000 ppm after

TABLE 8-56. CAPITAL AND OPERATING COST FOR LOW PRESSURE OXIDATION AND FILTRATION OF CONVENTIONAL BIOLOGICAL OXIDATION SLUDGES

Stream 4: Waste sludge from liquefaction biological oxidation system

Stream 5: Waste sludge from Lurgi gasification biological oxidation system

	Stream 4	Stream 5
Capacity, MGD	0.053	0.058
Estimated Capital Investment, $M	$750	$770
	Cost per day	Cost per day
Electric Power at $0.025/KWH	$26.75	$28.75
Cooling Water at $0.05/ 1,000 gals	0.77	0.83
Boiler Feed Water at $0.40/ 1,000 gals	7.44	8.04
Chemicals at $0.15/lb	1.20	1.20
Fuel at $0.45/gal	114.30	121.50
Maintenance Labor and Materials	45.00	45.00
Operating Labor at $80/man-shift	320.00	320.00
	$515.46	$525.32

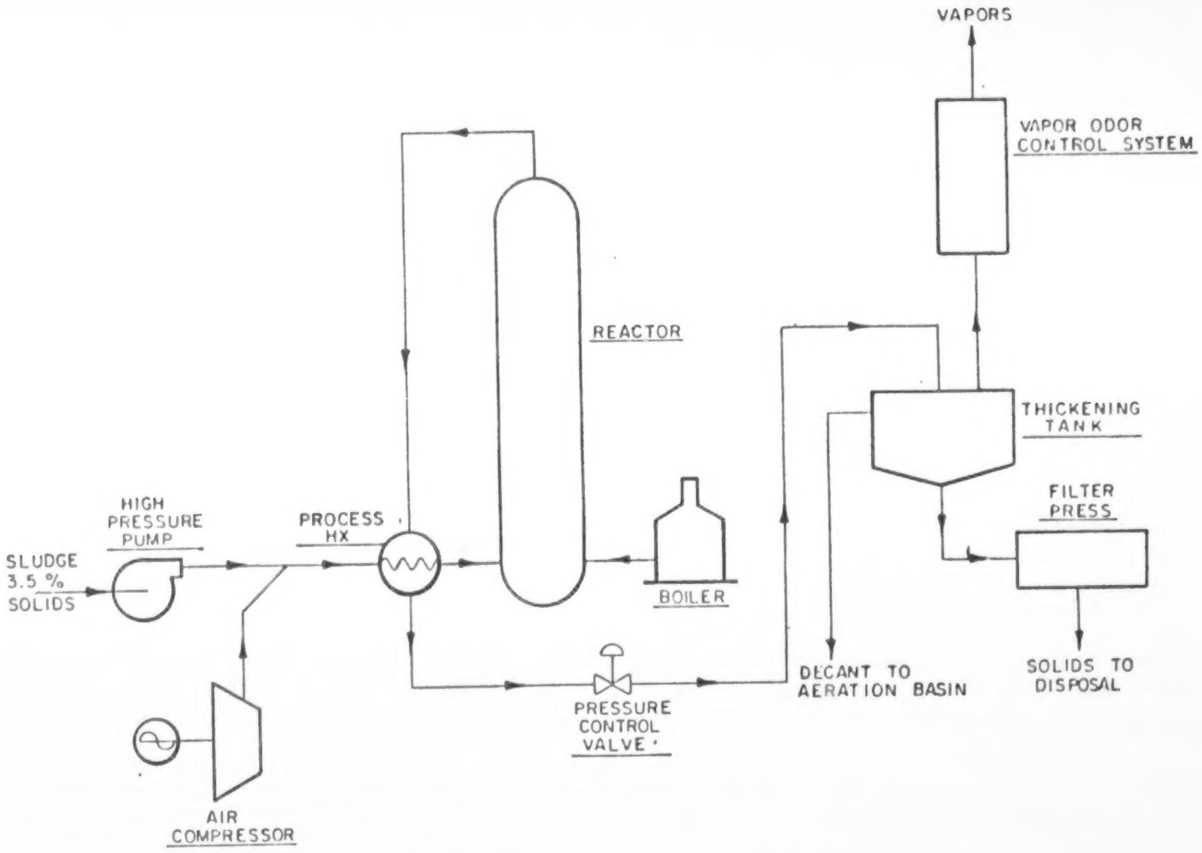

VAPORS
VAPOR ODOR CONTROL SYSTEM
REACTOR
THICKENING TANK
HIGH PRESSURE PUMP
PROCESS HX
FILTER PRESS
SLUDGE 3.5 % SOLIDS
BOILER
DECANT TO AERATION BASIN
SOLIDS TO DISPOSAL
PRESSURE CONTROL VALVE
AIR COMPRESSOR

Figure 8-49. Schematic of LPO & filter press for streams #4 & #5

TABLE 8-57. THE UNION CARBIDE "UNOX" SYSTEM IN WASTEWATER TREATMENT

Case I: Feed is wastewater from Lurgi gasification after phenol extraction, stripping, coagulation and flotation
Case III: Feed is wastewater from a liquefaction dissolver after treatment similar to Case I
(Quantities in mg/l unless otherwise stated)

	Most Stringent Standards (Pullman Kellogg Summary)	Case I Effluents		Case III Effluents		
		From UNOX Secondary Clarifier	From Multi-Media Filtration	From UNOX First Stage Clarifier	From UNOX Second Stage Clarifier	From Multi-Media Filtration
Flow, MGD		4.03	3.02	0.515	0.515	0.515
pH	6 to 9	7	7	7.5	7	7
TDS	1000	?	?	?	?	?
SS	15	30	15	30	30	<15
Sulfide	-	Insig(1)	Insig	Insig	Insig	Insig
Sulfate	600					
Nitrate	2.5	0	0			
Ammonia	2.5	1(2)	1(2)	1(2)	1(2)	1(2)
Cyanide	0.02	Insig	Insig	Insig	Insig	Insig
Thiocyanate	-	78	78	105	85	85
COD	125	?	?	?	?	?
BOD	30	>30	30	>455	>30	<30
Phenolics	0.005	0.0	0.1	<0.1	<0.1	<0.1
Oil and Grease	no visible	25	25	<25	<25	<25
Copper	0.1					
Mercury	0.002					

(1)Insig = Insignificant
(2)Ammonia added in slight excess

dilution by a receiving stream. The end of pipe discharge requirement for TDS concentrations is allowed to rise 750 ppm above background levels or up to 3,500 ppm if the additional TDS is caused by pollution abatement process or recycling. Furthermore, the effluent limitation for phenol is 0.3 ppm. Biological treatment can do an excellent job in removing phenolic material but present continuous monitoring devices cannot accurately monitor (much less detect) phenol concentrations down to 5 ppb. The COD effluent standard of 125 ppm requires tertiary treatment although an easily degradable waste such as the Case III waste (COD/BOD = 1.65) would have COD concentrations in that range. One EPA-funded study showed that biological treatment reduced the COD concentration of a coal conversion waste stream (COD/BOD = 1.5) from 6,500 ppm to 300 ppm. A more difficult waste like Case I (COD/BOD = 2.5) would have higher residual COD's but the effluent levels cannot be predicted without further data.

Design Description--The Case I wastewater is treated with a two-train, four-stage rectangular UNOX System having the dimensions of 208 ft by 104 ft by 14 ft plus 3 ft freeboard. Each stage contains one 75-NHp aerator to accomplish oxygen dissolution and solids mixing. The design is described and illustrated on the attached "quick estimate" documentation sheets and layout drawings. Comments regarding the process design are listed below:

1) Ammonia addition of 70 ppm is required.
2) Nitrification will not occur due to short sludge retention time (SRT) and elevated temperatures.
3) The cyanide and sulfide concentrations do not pose a toxicity problem.
4) The heavy metals will be adsorbed by the biomass and wasted from the system.

5) Thiocyanate will not be toxic but slow to degrade. About 60 to 80 percent will be removed.
6) An evaporative cooler is included in the system to maintain a biologically tolerable temperature below 104°F by removing 54 million Btu/hr from the internal recycle stream.

TABLE 8-57 shows the expected effluent quality after secondary clarification and multi-media filtration.

In Case III, a two-step UNOX System is incorporated to treat this high strength wastewater. The first step is a two-train, four-stage rectangular UNOX System having dimensions of 182 ft by 91 ft x 14 ft plus 3 ft freeboard. The first, second, and third stages contain 40 NHp aerators and the fourth stages contain a 50-NHp aerator for oxygen dissolution and solids mixing. Step 2 is a 2-MGD, one-train, two-stage rectangular modular UNOX System containing a 20-NHp aerator in each stage. Its dimensions are 59 ft by 29.5 ft by 10 ft plus 4 ft freeboard. The attached "quick estimate" documentation sheets describe both designs. As in Case I, the same process design comments apply. An ammonia addition of 425 ppm is required and, similarly, an evaporative cooler is included to maintain biologically tolerable temperatures in the reactor by removing 15 million Btu/hr from the internal recycle stream.

TABLE 8-57 shows the expected effluent quality coming after the first-step UNOX clarifiers, the second-step UNOX clarifiers and the multi-media filter.

General Comments--Further reduction in COD levels can be accomplished with ozonation. However, more information regarding COD characterization is required to determine an ozone dose.

Dilution water for Case III was available but not used because addition of that stream would not have simplified the UNOX System design nor reduced the amount of internal recycling for cooling purposes.

The UNOX System for Case I has been priced at $550,000 plus $400,000 for the required evaporative coolers. The Case III two-step UNOX System is priced at $550,000 plus $150,000 for the evaporative cooler.

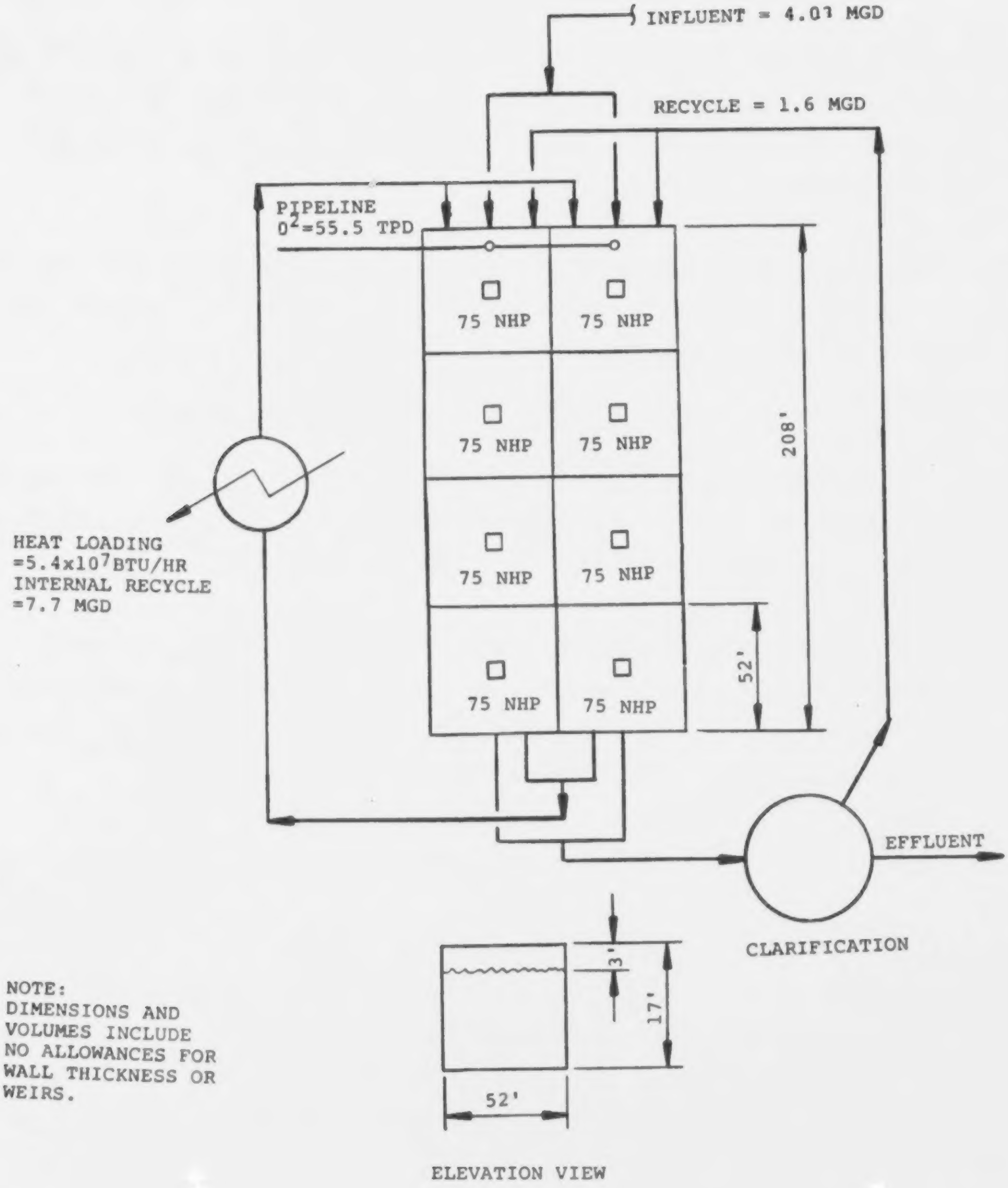
Union Carbide explained that prices quoted were only for the "UCC scope of supply" which is basically the aerators, purge blower and some instrumentation. Clarifier, reactor, pumps, etc. are not included.

"Quick Estimate" documentation sheets supplied by UCC are included as TABLES 8-58 and 8-59. Illustrative diagrams supplied for Cases I and III are presented as Figures 8-50 and 8-51.

In order to arrive at capital cost for the two UNOX Systems, for comparison with conventional air activated sludge, we applied the factor determined in the Pullman Kellogg estimate (i.e. Investment = purchased equipment ÷ 0.112):

Case I	$8,500,000	(Lurgi)
Case III	$6,250,000	(Liquefaction)

Phenosolvan Cost Information--
In response to telephone conversations and correspondence, American Lurgi furnished by telephone data on use of their proprietary Phenosolvan process for extraction of phenols from wastewater using isopropyl ether as the solvent. Figure 8-52 is a sketch illustrating the system.


INFLUENT = 4.03 MGD
RECYCLE = 1.6 MGD
PIPELINE
O2=55.5 TPD
75 NHP
75 NHP
75 NHP
75 NHP
75 NHP
75 NHP
75 NHP
75 NHP
208'
52'
HEAT LOADING
=5.4x10⁷BTU/HR
INTERNAL RECYCLE
=7.7 MGD
EFFLUENT
CLARIFICATION
3'
17'
52'
ELEVATION VIEW

NOTE:
DIMENSIONS AND
VOLUMES INCLUDE
NO ALLOWANCES FOR
WALL THICKNESS OR
WEIRS.

Figure 8-50.- Typical UNOX System layout for Case I.

TABLE 8-58. UNOX SYSTEM
"QUICK ESTIMATE" DOCUMENTATION

Customer: Pullman Kellogg
Location: Hypothetical
Consultant: Case I
Date: April 27, 1978
Sales Rep. R. W. Oeben
Engineer D. L. Wang

Design Basis:

Flow, MGD	4.03
Influent Wastewater Temp., °F	130
BOD , mg/l	2,000
BOD , lb/day	67,000
COD, mg/l	5,300
COD, lb/day	178,000
COD/BOD	2.65
NH , mg/l	30
Phenols, mg/l	200
Fatty acids, mg/l	560
Suspended Solids, mg/l	20
pH	7-9
VSS/TSS inf.	-

UNOX DESIGN:

Retention Time (on Q), hr	13.3
Biomass Loading, lb BOD /lb MLVSS-day	0.8
MLVSS Conc., mg/l	4,500
RSS Conc., %	2
Clarifier Overflow Rate, gal/ft -day	600
Oxygen Supply, tpd/Utilization, %	55.5/75
Dissolution: Power, BHP	562
, NHP/(Stagewise)	75/75/75/75
: Type	-
Oxygen Generation: Compressor Power, BHP/NHP Pipeline	99.5% 0 Purity

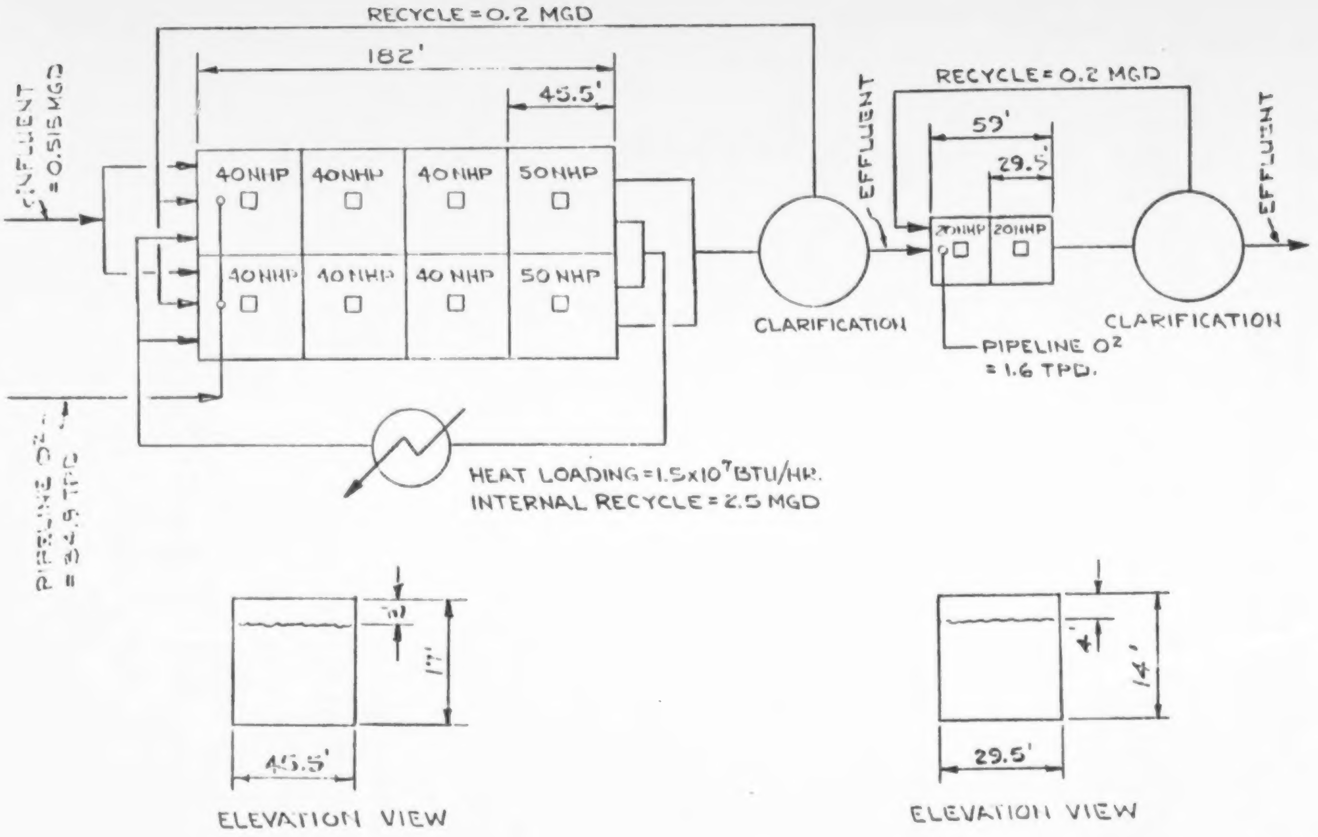

RECYCLE = 0.2 MGD
182'
45.5'
INFLUENT = 0.515 MGD
40 NHP
40 NHP
40 NHP
50 NHP
40 NHP
40 NHP
40 NHP
50 NHP
PIPELINE O2 = 34.9 TPD
HEAT LOADING = 1.5x10⁷ BTU/HR.
INTERNAL RECYCLE = 2.5 MGD
CLARIFICATION
EFFLUENT
RECYCLE = 0.2 MGD
59'
29.5'
20 NHP
20 NHP
PIPELINE O2 = 1.6 TPD.
CLARIFICATION
EFFLUENT
3'
17'
45.5'
ELEVATION VIEW
4'
14'
29.5'
ELEVATION VIEW

Figure 8-51. Typical UNOX System layout for Case III.

TABLE 8-59. "QUICK ESTIMATE" DOCUMENTATION

Customer:	Pullman Kellogg	Date:	April 27, 1978
Location:	Hypothetical	Sales Rep.	R. W. Oeben
Consultant:	Case III	Engineer	D. L. Wang

Design Basis:	First Step	Second Step
Flow, MGD	0.515	0.515
Influent Wastewater Temp., °F	130	102
BOD Load, lb/day	39,000	1,954
BOD Concentration, mg/l	9,100	455
COD/BOD	1.5	-
Suspended Solids mg/l	20	30
UNOX Design:		
Retention Time (on Q), hr	97	3.5
Biomass Loading, lb BOD /lb MLVSS-day	0.6	0.4
MLVSS Conc., mg/l	4,500	4,500
RSS Conc., %	2	2
Clarifier Overflow Rate, gal/ft day	600	600
Oxygen Supply, tpd/Utilization, %	34.9/70	1.6/75
Dissolution: Power, BHP,	312	38
NHP/(Stagewise)	40/40/40/50	20/20
: Type	UA	
Oxygen Generation:		
Compressor Power, BHP/NHP	-	-
: Size/Type	Pipeline O_2	
UNOX Reactor:		
Type	Rectangular	2 MGD Rect. Modular
Number of Trains/Stages	2/4	1/2
Overall Dimensions	182 x 91 x 14	59 x 29.5 x 10
Stage Dimensions	45.5 x 45.5 x 14	29.5 x 29.5 x 10
UNOX System Price: Oxygen Generation		
: UNOX System		
TOTAL		

Comments: Require 95% BOD_R in 1st and 2nd steps. Hypothetical coal conversion wastes. Estimate & Design will be used in EPA-funded document. System requires evaporative cooler to maintain biologically tolerable temperatures.

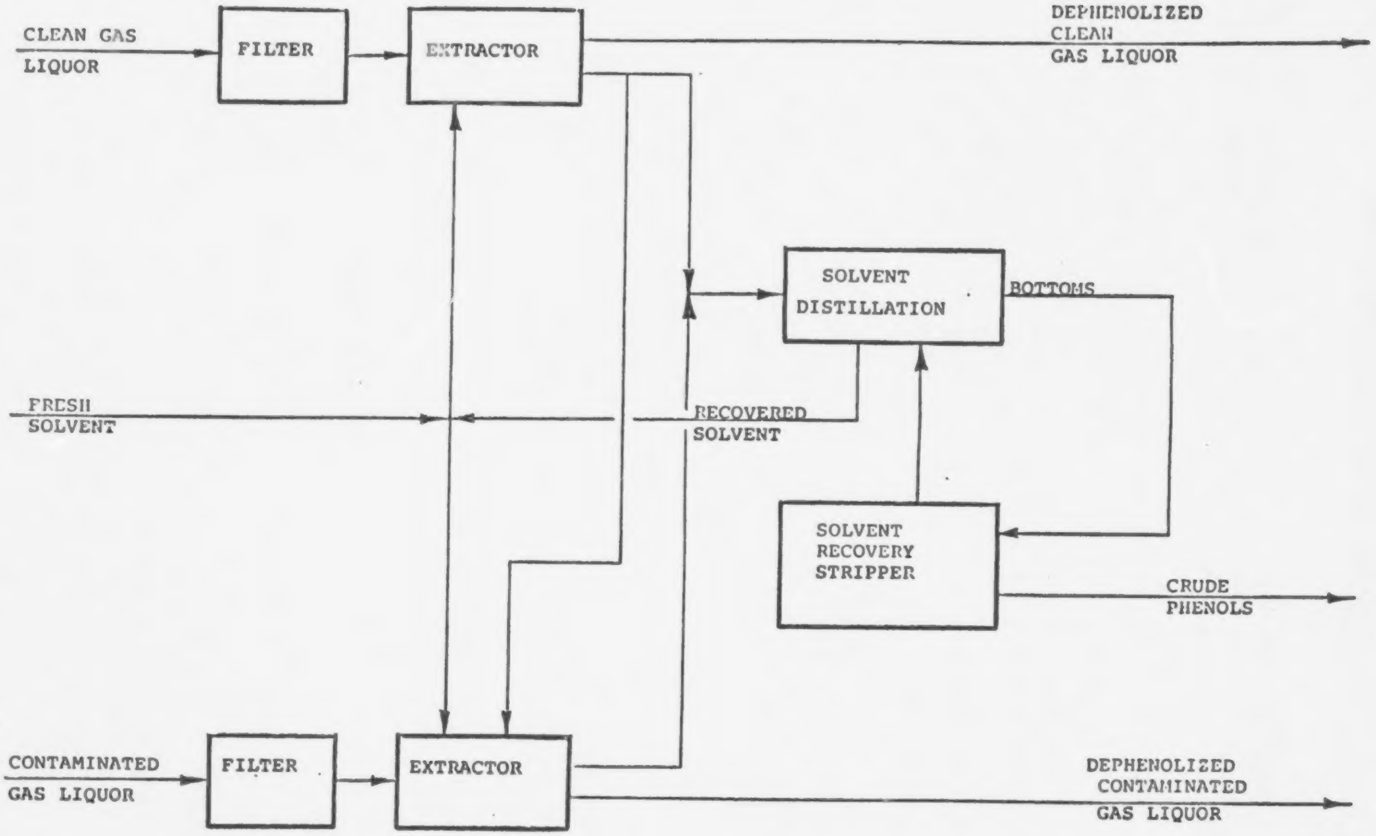

CLEAN GAS
LIQUOR
FILTER
EXTRACTOR
DEPHENOLIZED
CLEAN
GAS LIQUOR
SOLVENT
DISTILLATION
BOTTOMS
FRESH
SOLVENT
RECOVERED
SOLVENT
SOLVENT
RECOVERY
STRIPPER
CRUDE
PHENOLS
CONTAMINATED
GAS LIQUOR
FILTER
EXTRACTOR
DEPHENOLIZED
CONTAMINATED
GAS LIQUOR

Figure 8-52. Phenosolvan process.

(From Item 2 in reference list)

Flowrates and estimated compositions of two streams were furnished to Lurgi. Stream 1 was 185,000 lbs/hr of raw liquefaction condensate based on H-Coal data. Stream 2 was 1,400,000 lbs/hr of composition similar to Lurgi gasification or other p/o/t-producing processes.

Total installed costs were quoted as follows for Germany. American Lurgi believe U. S. Gulf Coast prices could be 30 to 40 percent higher.

	Stream 1		Stream 2
Wastewater, M lbs/hr	185		1,400
Capital Cost, DM	8,000,000	Single train	22,000,000
		Two train	30,000,000
U. S. Dollars, at $0.48/DM	3,900,000		10,670,000
		to	14,540,000
Escalate 35% to U.S. Gulf Coast	5,265,000		14,404,000
		to	19,630,000

It should be noted that the $12,000,000 estimated by Water Purification Associates (480) was reasonably close, considering the estimate accuracies with which we are dealing. Lurgi stated that the accuracy of their budget estimates is ± 25 percent, which is the general figure used by most contractors in quoting quick estimates.

Utility requirements quoted by Lurgi were:

Per 1,000 U. S. gallons of treated effluent	
Steam	80-270 lbs
Solvent Makeup	0.6-1.2 lbs
Cooling Water	1500 gal
Power	4-6 KWH

Raw Water Treatment for Removal of Inorganics--

L*A/Water Treatment Division furnished us with equipment costs of the usual processes involved in removing inorganics from raw water. Basis given to them was the raw water analysis used by C. F. Braun in their Lurgi process design on western coal (294, 295, 296). Costs supplied are shown in TABLE 8-60. L*A/WTD supplied us with an alternate set of figures, using the same treated water requirements, but assuming a very brackish raw water supply such as might be found in New Mexico.

Process	Through-put GPM	Equipment Cost
Pressure filters and	8 at 350 each	
Reverse Osmosis	2.016	$1,354,000
Demineralizer Trains	2 at 1.000 each	404,000
Condensate Polisher	730	85,000
H.P. BFW Deaerator	5,800	170,000
L.P. BFW Deaerator	840	35,000
Total Equipment		$2,048,000
Add-On as in TABLE 8-60 at 85%		1,740,800
Total Estimated Investment		$3,788,800

In both estimates the demineralizer trains included cation units and anion units in series.

TABLE 8-60. ESTIMATED CAPITAL INVESTMENT FOR REMOVAL OF INORGANICS FROM RAW WATER

Process	Throughput, GPM	Equipment Cost
Cold Lime Softener	1,620	
Gravity filters	3 at 750 each	
Clearwell	1,620	$ 291,680*
Zeolite Softeners	3 at 450 each	130,270*
Demineralizer Trains	2 at 1,000 each	404,000*
Ion Exchange Condensate Polisher	730	85,000*
H.P. BFW Deaerator	5,800	170,000
L.P. BFW Deaerator	840	35,000
Total Equipment		$1,115,950
Installation (estimated at 35%)		390,600
Instrumentation		Included
Piping (estimated at 20%)		223,000
Electrical (estimated at 10%)		111,600
Other indirects (estimated at 20%)		223,000
Total Estimated Investment		$2,064,150

---

*Total cost of equipment in the process

The add-ons are purely an "educated guess" based mostly on the Bechtel reference cited earlier. Freight is not included, and it is assumed that no buildings are necessary. In cold or rainy climates the assumption of no buildings would probably not be viable, but rough steel buildings or shelters would probably be considered acceptable in most cases.

Chemicals usages were cited, and we have applied current costs as follows:

Equipment	Chemical	Lbs/Day	¢/Lb	$/Day
Demineralizers	NaOH	6,132	17.5	$1,073
	H_2SO_4	6,132	2.5	153
Zeolite softener	NaCl	2,734	1.5	40
Cold lime softener	Alum, Lime and Polyelectrolyte		1.5	
Condensate polisher	NaOH	36.4	17.5	$ 6.37
	H_2SO_4	36.4	2.5	$ 0.91

Sludge quantities produced were stated, the largest of which is 25 percent of the feed to reverse osmosis as reject. In that case evaporation would probably be justified to concentrate the waste further and conserve water.

L*A/WTD offered some opinions on treatment required to use wastewater as a substitute for raw water as a boiler feed water and supplied information on the capability of reverse osmosis in removal of phenols, boron, ammonia, nitrates, Si, Ba,and Sr that was relayed from DuPont, the supplier of the membranes.

At our request, L*A/Water Treatment furnished additional cost information concerning the variation with capacity of the cost of cold lime clarifier, sodium zeolite softeners, reverse osmosis, and demineralizers in the range of the base case water quantities and the alternates.

A curve for budget cost selling price of reverse osmosis units is shown in Figure 8-53. The curve flattens at about 1 MGD, suggesting that duplicate units will be required for higher capacities.

Other comments by L*A/Water Treatment were as follows:

- For the cold lime clarifier, cost for capacities above the 809,000 lbs/hr rate on which the quotation was based would have to be assumed to be directly proportional to flow, for budget purposes, although resulting estimated costs will probably be high.

- For sodium zeolite softeners, costs for capacity above the original 436,000 lbs/hr may be estimated at $150/gpm for the additional throughput.

- For ion exchange units, costs for capacity above the original 569,000 lb/hr would be directly proportional to flow.

- In the absence of tests, it is generally assumed that reverse osmosis will reject at least 90 percent of the TDS if a 25 percent reject (loss) stream is assumed. Temperature of operation should be restricted to 65 to 95° F so that membrane life may be guaranteed and operation will be satisfactory.

- Recycling Lurgi base case treated water after post-biox

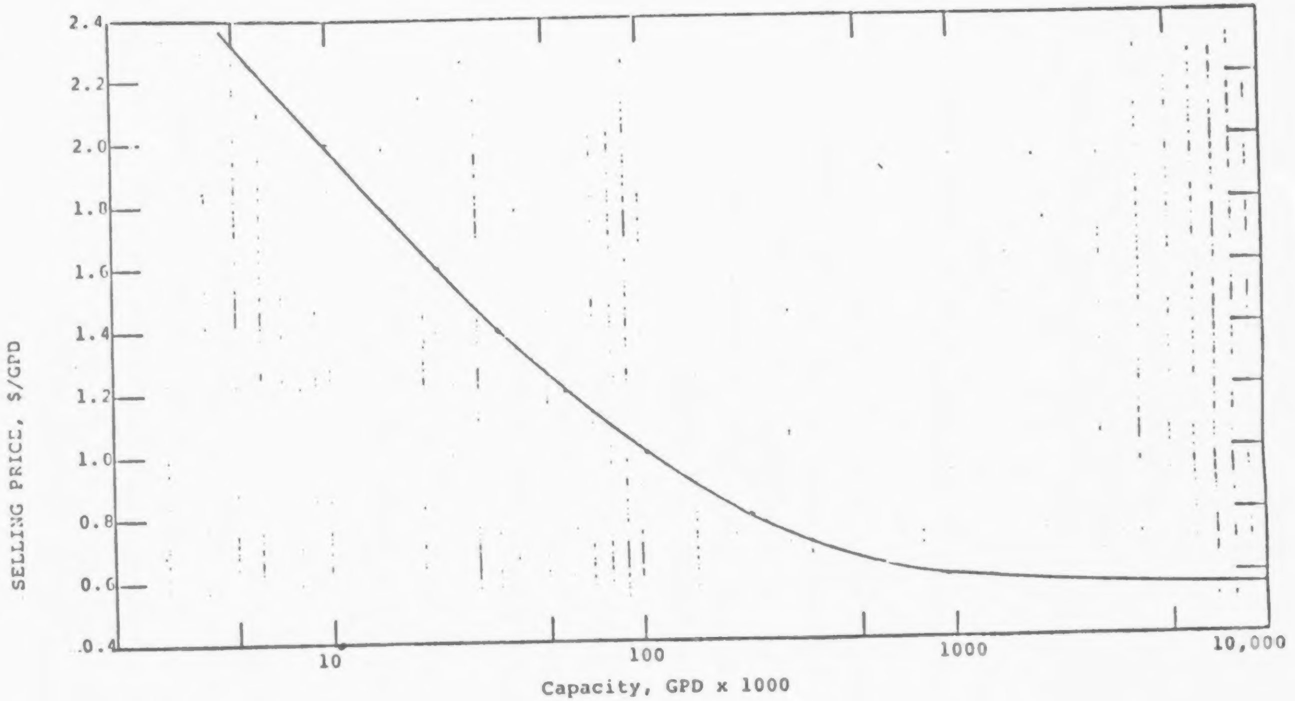


Figure 8-53. Reverse osmosis: budget prices without pretreatment (add 15% for complete pretreatment system.*)

*Communication from L*A Water Treatment

filtration and reverse osmosis followed by ion exchange should be feasible. Tests to check the effect on membrane fouling of the small (1 mg/l) oil and grease residual were recommended. It was stated that a periodic cleaning with detergent (Biz) might be needed.

o Direct ion exchange of stripped wastewater from gasification processes producing no p/o/t might require some pretreatment to reduce the estimated 52 mg/l COD. This level of COD usually indicates to the vendor that a possible bacteria fouling problem exists. Chlorination and carbon filtration or the use of ultraviolet sterilizers are normally recommended if this is found to be the case. The problem seems to be unlikely unless the wastewater has been exposed to open air conditions. Operating costs for ion exchange were cited as 3¢/lb H_2SO_4, 8¢/lb NaOH, and 50¢/100 gals of wastewater.

o We included a query as to whether reverse osmosis could substitute for evaporation of inorganic sludges (see Lurgi base case 1 Braun design in the discussion of evaporator costs) such as demineralizer waste, cooling tower blowdown, and sludges from the zeolite softener and the lime clarifier. The answer was that estimated TDS entering reverse osmosis would be 5,000 ppm, that the maximum recovery would probably drop from the classic 75 percent to 50 to 60 percent and that 500 to 600 ppm would pass the membrane to be removed by ion exchange. Reject flow of 40 to 50 percent would then go to the evaporator as would the additional demineralizer waste.

Evaporator Costs--

Two similar cases representing the quantities and streams evaporated in the Braun conceptual designs for Lurgi (p/o/t produced) and Bi-Gas (no p/o/t produced) were submitted to Envirotech for design and cost information. Envirotech obtained the following information from their Goslin group which specializes in evaporation.

Goslin estimates are based on a forced circulation, six-effect evaporator-crystallizer. Although steam economy could be improved by incorporation of a feed preheater, no preheater was included due to their opinion that $CaSO_4$ scaling would probably be a problem. The feed compositions for the two cases, which we believe to be somewhat high, were transmitted to Goslin as follows:

	Case 1	Case 2
Cooling Tower Blowdown	69,500 lbs/hr	74,300 lbs/hr
Sodium Softener Sludge	146,000	76,740
Condensate Polisher Waste	4,000	10,280
Cold Lime Clarifier Sludge	64,200	46,830
Demineralizer Regeneration Waste	-	107,430
	283,700 lbs/hr	315,580 lbs/hr

TABLE 8-61 presents the design and cost information. Note that installed costs for Case 1 (Lurgi) are estimated at $4,000,000 and for Case 2 (Bi-Gas) at $4,700,000 for 283,700 and 315,580 lbs/hr, respectively, of evaporator feed.

TABLE 8-61. PULLMAN KELLOGG GASIFICATION STUDY: EVAPORATOR COSTS

EVAPORATION	Case 1	Case 2
Feedrate, lbs/hr	283,700	315,580
Steam press, psig	25	18
Total requirement, lbs/hr	61,000	67,800
Goslin Evaporator Size		
Vapor Heads, Dia.	11-1/2 to 14-1/2'	12 to 15'
Heating elements, ft^2	11,260	14,340
Pumps, HP	100	125
Condenser	Surface	Surface
Cooling Water, GPM (1)	2,100	2,350
Ejector - 2 stage		
Steam, lbs/hr	300	300
Cooling water, GPM	20	20
Steam economy (overall)	4.46	4.5
Total weight, lbs.	520,000	600,000
Total cost, equipment (2)	$2,000,000	$2,350,000
Installed factor (3)	2.0	2.0
Total Installed Cost	$4,000,000	$4,700,000

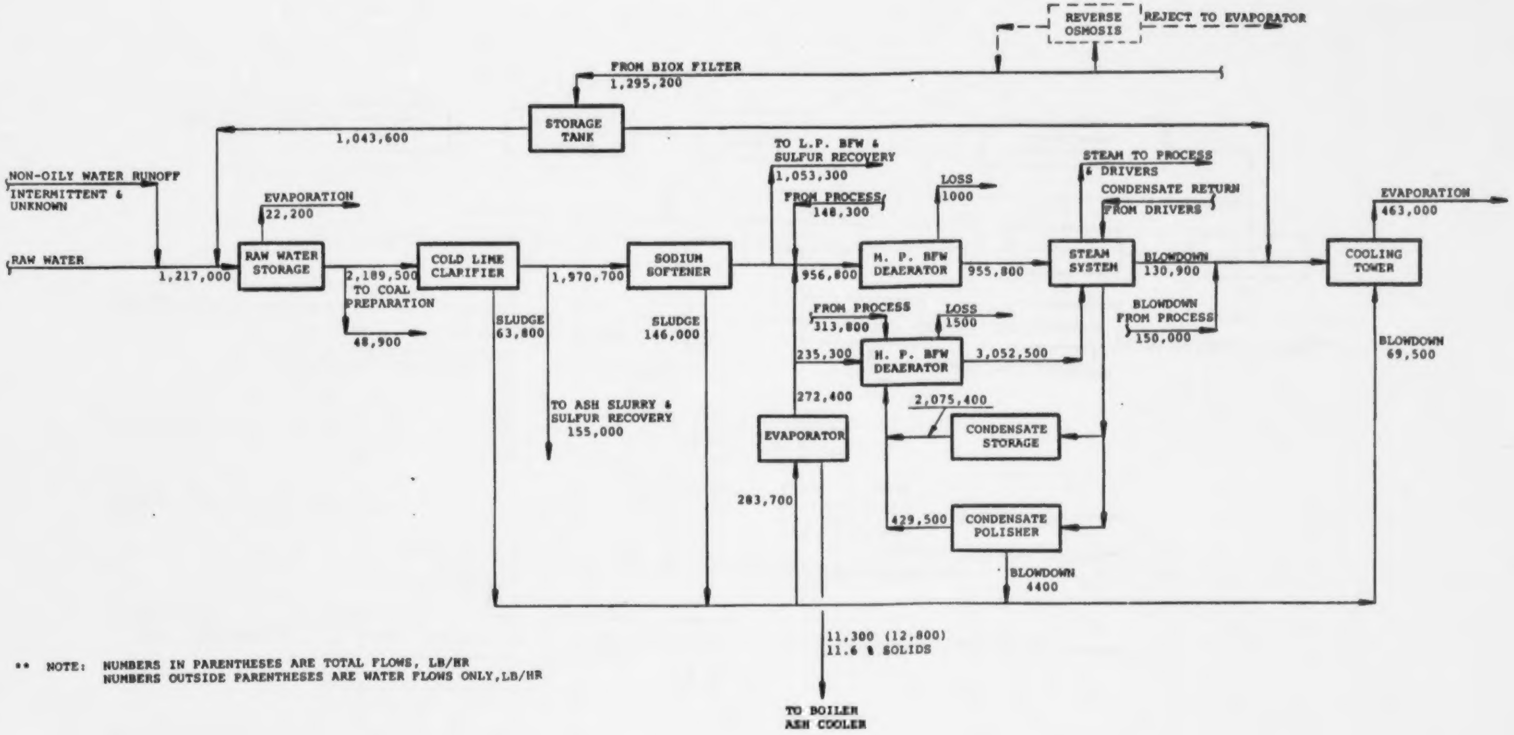
1. Cooling water - assumed 90° F temperature.
2. Price is FOB Birmingham, Alabama. Based on 316 SS, backward feed and includes six large diameter vapor pipes, six recirculating pumps and piping.
3. Installed cost is double the equipment cost and includes Goslin six effect forced circulation evaporator with instrumentation, small piping, foundation, structure, insulation, wiring, paint, site work and engineering.

INTEGRATED SCHEMES FOR WASTEWATER TREATMENT

As previously mentioned, three base cases from conceptual designs have been selected to obtain water quantities for study of water treatment methods and sludge production and disposal from these. Figure 8-54 is a block flow diagram illustrating the entire water system and treating processes employed by C. F. Braun for Lurgi (p/o/t-producing) gasification. Estimated analyses of key streams are tabulated in the lower portion of the figure. The sour water analyses were obtained from sources other than the C. F. Braun report. Results of treating the sour water are Pullman Kellogg's best estimates and must be confirmed by testing. Various alternates to the Braun treating scheme for "zero discharge" will be discussed and illustrated by sketches or tabulations.

Figure 8-55 is a similar block flow diagram illustrating a high temperature entrained flow gasification process that produces no p/o/t. Quantities shown are from the conceptual design for Bi-Gas by C. F. Braun. The sour water analysis was taken from Kopper-Totzek data, since Bi-Gas analyses and treating data have not yet been published. Various alternates to the Braun treating scheme will be considered. Results of treating methods are Pullman Kellogg's best estimates and must be confirmed.

Figure 8-56 is the water system base case for an integrated liquefaction process. Quantities shown are from the Ralph M. Parsons conceptual design for SRC II, as previously mentioned. In Parsons' design the stripped sour water was returned to the process gasifier, which produces the necessary hydrogen for liquefaction, and all organic materials were assumed to be destroyed by combustion at the high temperature of the Bi-Gas reactor. This may well be the preferred treatment, but it should be demonstrated in an integrated pilot or demonstration unit. We believe the organic compounds will indeed be destroyed, but


REVERSE OSMOSIS
REJECT TO EVAPORATOR
FROM BIOX FILTER
1,295,200
STORAGE TANK
1,043,600
NON-OILY WATER RUNOFF
INTERMITTENT & UNKNOWN
RAW WATER
1,217,000
EVAPORATION
22,200
RAW WATER STORAGE
2,189,500
TO COAL PREPARATION
48,900
COLD LIME CLARIFIER
SLUDGE
63,800
1,970,700
TO ASH SLURRY & SULFUR RECOVERY
155,000
SODIUM SOFTENER
SLUDGE
146,000
TO L.P. BFW & SULFUR RECOVERY
1,053,300
FROM PROCESS
148,300
956,800
M. P. BFW DEAERATOR
LOSS
1000
955,800
FROM PROCESS
313,800
LOSS
1500
235,300
H. P. BFW DEAERATOR
3,052,500
272,400
EVAPORATOR
283,700
2,075,400
CONDENSATE STORAGE
429,500
CONDENSATE POLISHER
BLOWDOWN
4400
STEAM TO PROCESS & DRIVERS
CONDENSATE RETURN FROM DRIVERS
STEAM SYSTEM
BLOWDOWN
130,900
BLOWDOWN FROM PROCESS
150,000
EVAPORATION
463,000
COOLING TOWER
BLOWDOWN
69,500
11,300 (12,800)
11.6 % SOLIDS
TO BOILER ASH COOLER

** NOTE: NUMBERS IN PARENTHESES ARE TOTAL FLOWS, LB/HR
NUMBERS OUTSIDE PARENTHESES ARE WATER FLOWS ONLY, LB/HR

Figure 8-54a.- Integrated scheme for treatment of Lurgi wastewaters.

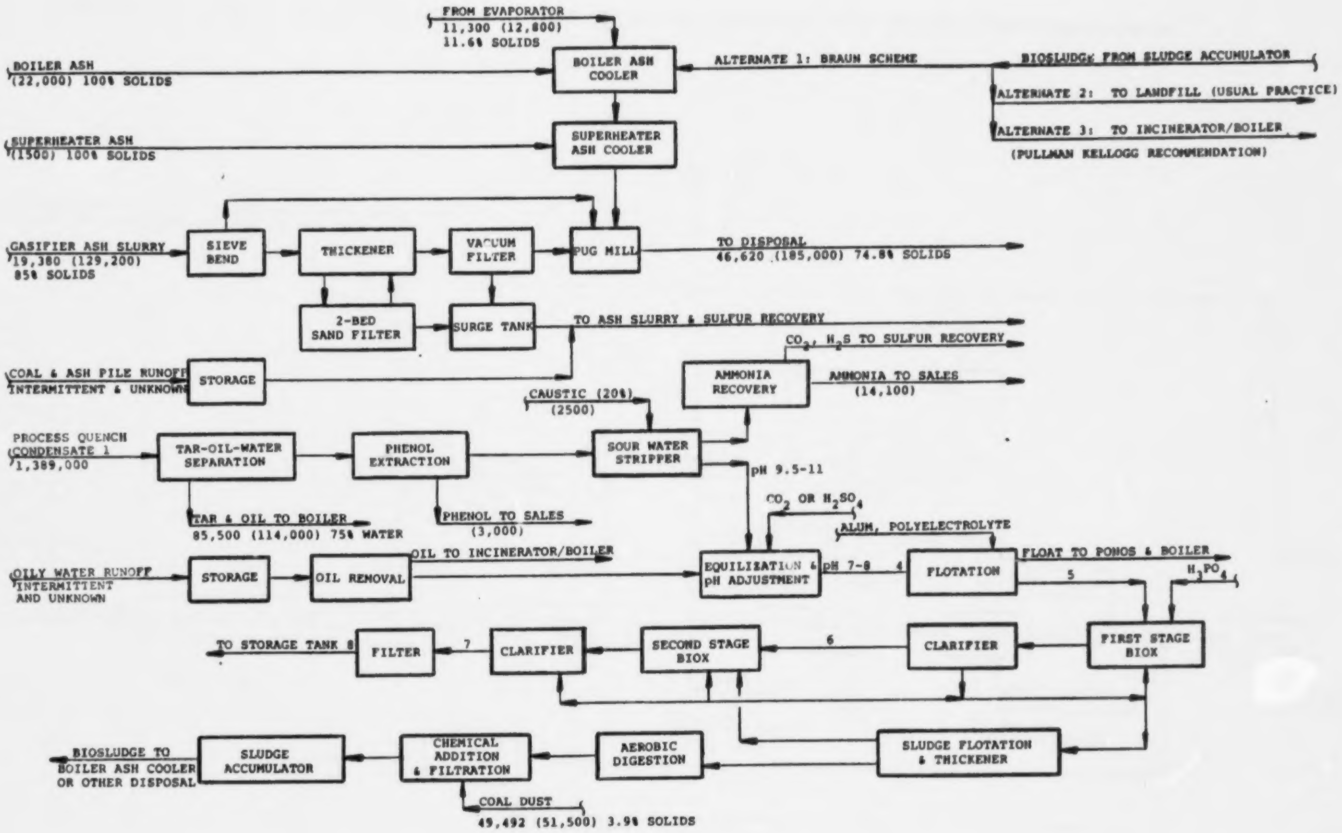
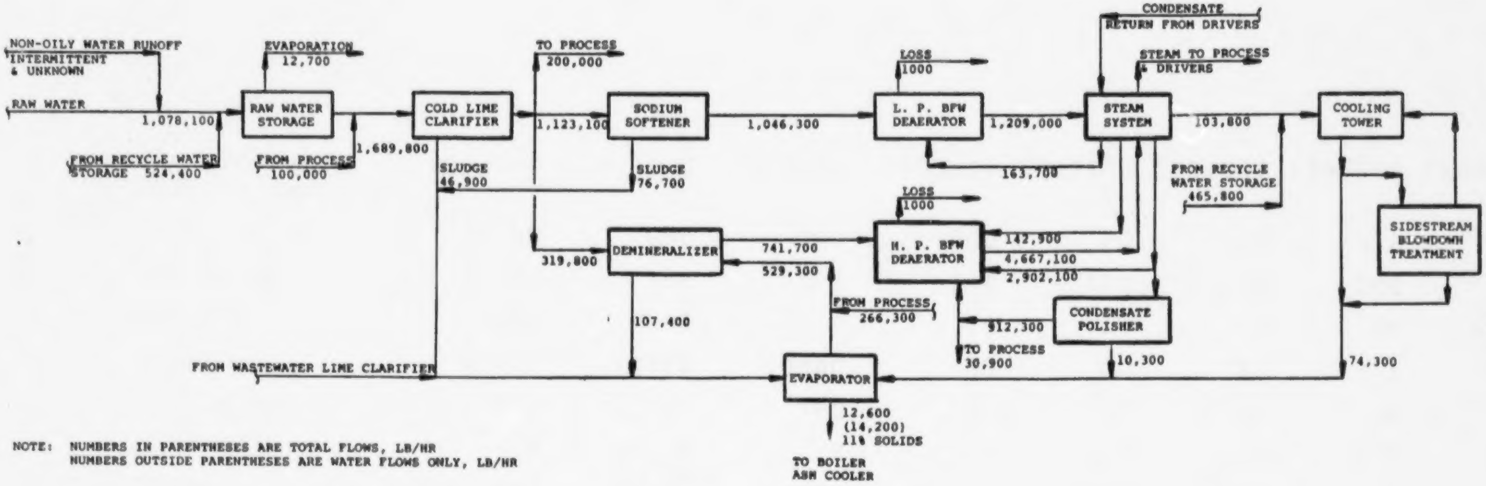

FROM EVAPORATOR
11,300 (12,800)
11.6% SOLIDS
BOILER ASH
(22,000) 100% SOLIDS
BOILER ASH COOLER
ALTERNATE 1: BRAUN SCHEME
BIOSLUDGE FROM SLUDGE ACCUMULATOR
ALTERNATE 2: TO LANDFILL (USUAL PRACTICE)
ALTERNATE 3: TO INCINERATOR/BOILER
(PULLMAN KELLOGG RECOMMENDATION)
SUPERHEATER ASH
(1500) 100% SOLIDS
SUPERHEATER ASH COOLER
GASIFIER ASH SLURRY
19,380 (129,200)
85% SOLIDS
SIEVE BEND
THICKENER
VACUUM FILTER
PUG MILL
TO DISPOSAL
46,620 (185,000) 74.8% SOLIDS
2-BED SAND FILTER
SURGE TANK
TO ASH SLURRY & SULFUR RECOVERY
CO2, H2S TO SULFUR RECOVERY
COAL & ASH PILE RUNOFF
INTERMITTENT & UNKNOWN
STORAGE
AMMONIA RECOVERY
AMMONIA TO SALES
(14,100)
CAUSTIC (20%)
(2500)
PROCESS QUENCH
CONDENSATE 1
1,389,000
TAR-OIL-WATER SEPARATION
PHENOL EXTRACTION
SOUR WATER STRIPPER
pH 9.5-11
TAR & OIL TO BOILER
85,500 (114,000) 75% WATER
PHENOL TO SALES
(3,000)
CO2 OR H2SO4
ALUM, POLYELECTROLYTE
OIL TO INCINERATOR/BOILER
OILY WATER RUNOFF
INTERMITTENT
AND UNKNOWN
STORAGE
OIL REMOVAL
EQUILIZATION & pH ADJUSTMENT
pH 7-8
4
FLOTATION
FLOAT TO PONDS & BOILER
5
H3PO4
TO STORAGE TANK 8
FILTER
7
CLARIFIER
SECOND STAGE BIOX
6
CLARIFIER
FIRST STAGE BIOX
BIOSLUDGE TO
BOILER ASH COOLER
OR OTHER DISPOSAL
SLUDGE ACCUMULATOR
CHEMICAL ADDITION & FILTRATION
AEROBIC DIGESTION
SLUDGE FLOTATION & THICKENER
COAL DUST
49,492 (51,500) 3.9% SOLIDS

Figure 8-54b.- Integrated scheme for treatment of Lurgi wastewaters.

Streams in Figure 8-54

	1	2	3	4	5	6	7	8
	(mg/l unless otherwise noted)							
BOD	7,200	6,000	2,320	2,070	1,700	170	17	8
COD	13,000	10,400	6,220	5,220	4,650	1,000	500	400
TDS	1,884	1,884	1,884					
TSS	4,676	468	468	125	30	20	20	5
Phenol	3,100	2,500	410	410	410	20	1	1
Cyanide	8	8	8	5	5	2	0.6	0.6
Thiocyanate	260	260	260	260	260	50	5	5
Ammonia	13,600	10,900	10,900	80	80	30	1	0.9
Sulfide	506	506	506	10	10	2	0.06	0.06
Chloride	266	266	266	266	266	266	266	266
Oil	21,000	500	500	500	50	10	5	1
Water flow, lb/hr	1,389,000	1,303,600	1,303,600	1,303,600	1,247,700	1,129,300	1,295,200	1,295,200

Figure 8-54(c). Stream compositions for integrated scheme for treatment of Lurgi wastewaters (part 3 of 3).


NON-OILY WATER RUNOFF
INTERMITTENT & UNKNOWN
RAW WATER
1,078,100
FROM RECYCLE WATER STORAGE 524,400
EVAPORATION 12,700
RAW WATER STORAGE
FROM PROCESS 100,000
1,689,800
COLD LIME CLARIFIER
SLUDGE 46,900
TO PROCESS 200,000
1,123,100
SODIUM SOFTENER
SLUDGE 76,700
319,800
DEMINERALIZER
107,400
1,046,300
741,700
529,300
FROM PROCESS 266,300
EVAPORATOR
12,600
(14,200)
11% SOLIDS
TO BOILER ASH COOLER
FROM WASTEWATER LIME CLARIFIER
LOSS 1000
L. P. BFW DEAERATOR
1,209,000
163,700
LOSS 1000
H. P. BFW DEAERATOR
142,900
4,667,100
2,902,100
912,300
TO PROCESS 30,900
CONDENSATE RETURN FROM DRIVERS
STEAM TO PROCESS & DRIVERS
STEAM SYSTEM
CONDENSATE POLISHER
10,300
103,800
FROM RECYCLE WATER STORAGE 465,800
COOLING TOWER
SIDESTREAM BLOWDOWN TREATMENT
74,300

NOTE: NUMBERS IN PARENTHESES ARE TOTAL FLOWS, LB/HR
NUMBERS OUTSIDE PARENTHESES ARE WATER FLOWS ONLY, LB/HR

Figure 8-55a.- Integrated scheme for treatment of wastewater from gasification processes producing no p/o/t.

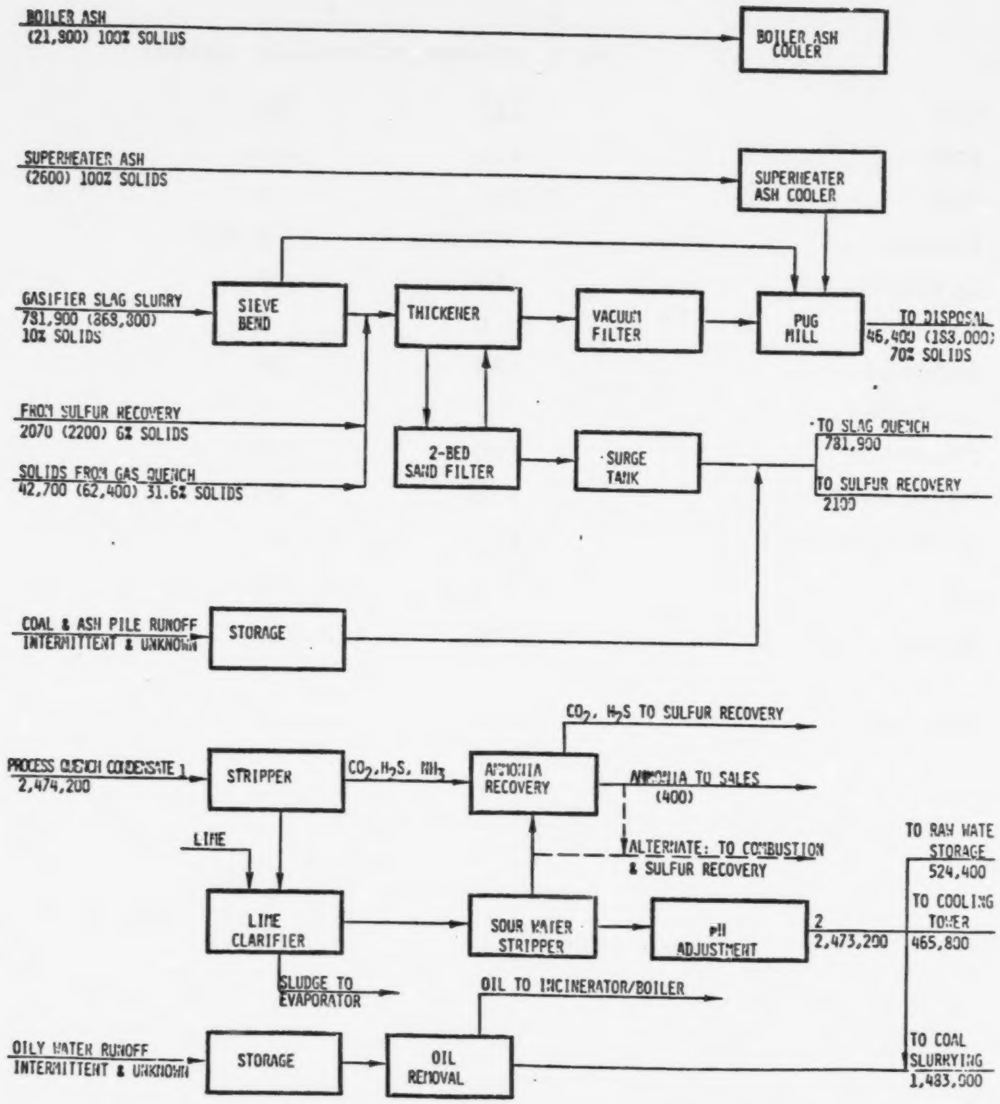

BOILER ASH
(21,800) 100% SOLIDS
BOILER ASH COOLER
SUPERHEATER ASH
(2600) 100% SOLIDS
SUPERHEATER ASH COOLER
GASIFIER SLAG SLURRY
781,900 (868,300)
10% SOLIDS
SIEVE BEND
THICKENER
VACUUM FILTER
PUG MILL
TO DISPOSAL
46,400 (153,000)
70% SOLIDS
FROM SULFUR RECOVERY
2070 (2200) 6% SOLIDS
SOLIDS FROM GAS QUENCH
42,700 (62,400) 31.6% SOLIDS
2-BED SAND FILTER
SURGE TANK
TO SLAG QUENCH
781,900
TO SULFUR RECOVERY
2100
COAL & ASH PILE RUNOFF
INTERMITTENT & UNKNOWN
STORAGE
CO2, H2S TO SULFUR RECOVERY
PROCESS QUENCH CONDENSATE 1
2,474,200
STRIPPER
CO2, H2S, NH3
AMMONIA RECOVERY
AMMONIA TO SALES
(400)
ALTERNATE: TO COMBUSTION
& SULFUR RECOVERY
LIME
LIME CLARIFIER
SOUR WATER STRIPPER
pH ADJUSTMENT
2
2,473,200
TO RAW WATE STORAGE
524,400
TO COOLING TOWER
465,800
SLUDGE TO EVAPORATOR
OIL TO INCINERATOR/BOILER
OILY WATER RUNOFF
INTERMITTENT & UNKNOWN
STORAGE
OIL REMOVAL
TO COAL SLURRYING
1,483,900

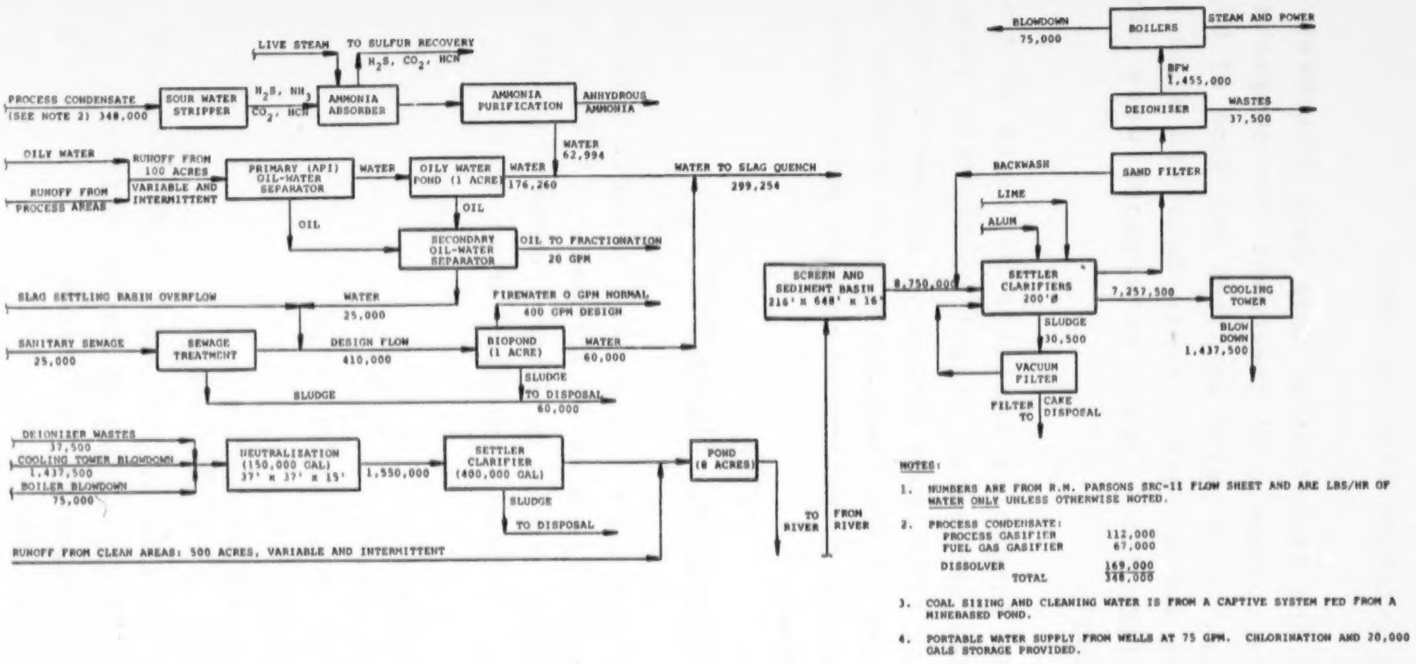
Figure 8-55(b). Integrated scheme for treatment of wastewater from gasification processes producing no p/o/t (part 2 of 3).

Note: Numbers in parentheses are total flows, lb/hr
Numbers outside parentheses are water flows only/ lb/hr

	Streams in Figure 8-55	
	1	2
	(mg/l unless otherwise noted)	
COD	128	52
TDS	831	475
TSS	5,081	5
Phenol	0.011	0.011
Cyanide	13	6
Oil	0	0
Ammonia	184	20
Sulfide	7	7
Chloride	100	100
Hardness*	630	80
Alkalinity*	650	10
pH	8.5	7
Water flow, lb/hr	2,474,200	2,473,200

*As $CaCO_3$

Figure 8-55(c). Integrated scheme for treatment of wastewaters from gasification processes producing no p/o/t (part 3 of 3).


PROCESS CONDENSATE (SEE NOTE 2) 348,000
SOUR WATER STRIPPER
H_2S, NH_3
CO_2, HCN
LIVE STEAM
TO SULFUR RECOVERY
H_2S, CO_2, HCN
AMMONIA ABSORBER
AMMONIA PURIFICATION
ANHYDROUS AMMONIA
WATER 62,994
OILY WATER
RUNOFF FROM PROCESS AREAS
RUNOFF FROM 100 ACRES
VARIABLE AND INTERMITTENT
PRIMARY (API) OIL-WATER SEPARATOR
WATER
OILY WATER POND (1 ACRE)
WATER 176,260
OIL
OIL
SECONDARY OIL-WATER SEPARATOR
OIL TO FRACTIONATION
20 GPM
WATER TO SLAG QUENCH
299,254
SLAG SETTLING BASIN OVERFLOW
WATER
25,000
FIREWATER 0 GPM NORMAL
400 GPM DESIGN
SANITARY SEWAGE
25,000
SEWAGE TREATMENT
DESIGN FLOW
410,000
BIOPOND (1 ACRE)
WATER
60,000
SLUDGE
SLUDGE
TO DISPOSAL
60,000
DEIONIZER WASTES
37,500
COOLING TOWER BLOWDOWN
1,437,500
BOILER BLOWDOWN
75,000
NEUTRALIZATION (150,000 GAL) 37' x 37' x 15'
1,550,000
SETTLER CLARIFIER (400,000 GAL)
SLUDGE
TO DISPOSAL
POND (8 ACRES)
RUNOFF FROM CLEAN AREAS: 500 ACRES, VARIABLE AND INTERMITTENT
TO RIVER
FROM RIVER
SCREEN AND SEDIMENT BASIN 216' x 648' x 16'
8,750,000
BLOWDOWN
75,000
BOILERS
STEAM AND POWER
BFW
1,455,000
DEIONIZER
WASTES
37,500
BACKWASH
SAND FILTER
LIME
ALUM
SETTLER CLARIFIERS 200'Ø
7,257,500
COOLING TOWER
BLOW DOWN
1,437,500
SLUDGE
30,500
VACUUM FILTER
FILTER CAKE TO DISPOSAL

NOTES:

1. NUMBERS ARE FROM R.M. PARSONS SRC-II FLOW SHEET AND ARE LBS/HR OF WATER ONLY UNLESS OTHERWISE NOTED.
2. PROCESS CONDENSATE:

PROCESS GASIFIER	112,000
FUEL GAS GASIFIER	67,000
DISSOLVER	169,000
TOTAL	348,000

3. COAL SIZING AND CLEANING WATER IS FROM A CAPTIVE SYSTEM FED FROM A MINEBASED POND.
4. PORTABLE WATER SUPPLY FROM WELLS AT 75 GPM. CHLORINATION AND 20,000 GALS STORAGE PROVIDED.

Figure 8-56.- Liquefaction base case water balance.

the fate of the inorganic compounds must be traced to confirm that they do not cause operational problems such as buildup, scaling, downstream catalyst fouling, etc.

In the alternate to the liquefaction base case flowsheet Pullman Kellogg shows a more conventional treatment for comparison with that used by Parsons. Sour water analyses and some of the treating results are from H-Coal bench scale data and from treating experiments by AWARE, Inc. These references were mentioned earlier in this report.

As described in the foregoing sections on costs of treatment, Pullman Kellogg has attempted to get up-to-date budget cost figures and treating efficiency advice or statements from various vendors and licensors of the treating modules shown on the flowsheets, and information on alternate methods which could be used instead of those shown.

Lurgi Gasification Flowsheet, Figure 8-54

This process produces tars, oils, and phenols and all water issuing from the gasifiers must be treated. It thus represents the most extensive and expensive water treatment of all the gasification processes. Treatment modules shown are:

- Oil Separation
- Phenol Extraction
- Stripping of CO_2, H_2S, and NH_3
- Ammonia Recovery
- Dissolved Air Flotation
- Equalization and pH Adjustment
- Biological Oxidation, including Sludge Handling
- Filtration of Biological Oxidation Effluent
- Evaporation of Inorganic Sludges

The thickened, stabilized sludge from biological oxidation was used for ash cooling in the Braun conceptual design and passed out with the collected ash and inorganic solids from the evaporator. Pullman Kellogg suggests that this organic material be burned in a central incinerator boiler with the tars and crude phenols which may not be sold (see Section 9, Control of Gaseous Emissions). A common method of disposal is in a sanitary landfill and this alternate is also indicated on the flowsheet. Problems associated with the sanitary landfill disposal method are discussed in Section 10. Using the costs discussed in the foregoing section, our best estimate of the capital costs for the wastewater treating scheme shown on the flowsheet are presented in TABLE 8-62.

Lurgi Gasification - Alternate Treating Schemes

One fairly simple alternate to the Braun treating scheme would be to reduce the quantity of the sludges to be fed to the four-stage evaporator by employing reverse osmosis, demineralization, and a small evaporator on the reject stream from reverse osmosis. This is illustrated in Figure 8-57.

Capital cost of the alternate case is estimated as follows:

Reverse Osmosis with Pre-filter	$ 632,000	(1)
Demineralizer Train	224,000	(1)
Total Equipment (except evaporator)	$ 856,000	
Installation Cost (0.8 x equipment)	684,000	(2)
Total Installed Cost	$1,540,000	
Evaporator Installed Cost	1,890,000	
Capital Cost	$3,430,000	

(1) Equipment cost furnished by L*A Water Treatment

(2) Factor from Bechtel paper "WateReuse - 1975."

TABLE 8-62. ESTIMATED CAPITAL COSTS FOR WASTEWATER TREATING

Item	Capacity, MGD	Capital, $MM	Source/ Reference
API Separator	3.8	0.55	SRI Report (804)
Phenol Extraction	3.7	14.40*	American Lurgi
Stripping and Ammonia Recovery	3.6	12.0	Chevron Research
Dissolved Air Flotation	3.6	Included in biological oxidation	Envirotech/Kellogg
Equalization Basin	(24 hr. hold time) 3.6	0.15	Pullman Kellogg
Pumping Station		0.39	Pullman Kellogg
Biological Oxidation**	3.6	10.00	Envirotech/Kellogg
Sludge Belt Press		0.47	H.O. Schultz
Evaporator	0.816	4.0	Goslin/Envirotech
Total Capital Cost		41.96	--

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* Single-train plant, quoted at $10,670,000 in Germany, escalated by 35 percent for U.S. Gulf Coast. Two-train plant quoted at $14,540,000 in Germany, escalates to $19.6 million for U.S. Gulf Cost.

**Includes effluent filters, sludge thickener, and aerobic sludge digestion.

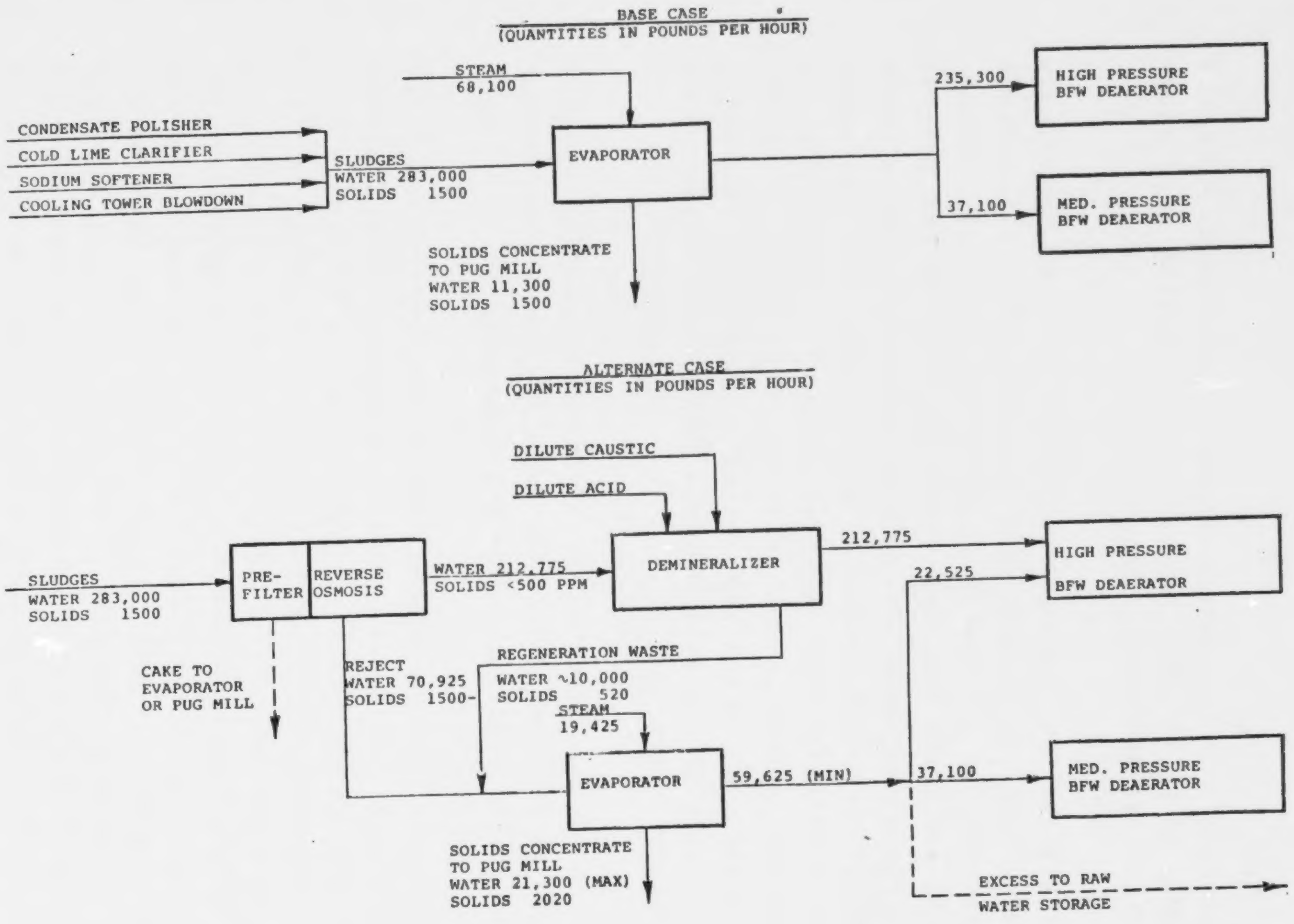
BASE CASE
(QUANTITIES IN POUNDS PER HOUR)
STEAM
68,100
CONDENSATE POLISHER
COLD LIME CLARIFIER
SODIUM SOFTENER
COOLING TOWER BLOWDOWN
SLUDGES
WATER 283,000
SOLIDS 1500
EVAPORATOR
SOLIDS CONCENTRATE
TO PUG MILL
WATER 11,300
SOLIDS 1500
235,300
HIGH PRESSURE
BFW DEAERATOR
37,100
MED. PRESSURE
BFW DEAERATOR
ALTERNATE CASE
(QUANTITIES IN POUNDS PER HOUR)
DILUTE CAUSTIC
DILUTE ACID
SLUDGES
WATER 283,000
SOLIDS 1500
PRE-
FILTER
REVERSE
OSMOSIS
WATER 212,775
SOLIDS <500 PPM
DEMINERALIZER
212,775
HIGH PRESSURE
BFW DEAERATOR
22,525
CAKE TO
EVAPORATOR
OR PUG MILL
REJECT
WATER 70,925
SOLIDS 1500-
REGENERATION WASTE
WATER ~10,000
SOLIDS 520
STEAM
19,425
EVAPORATOR
59,625 (MIN)
37,100
MED. PRESSURE
BFW DEAERATOR
SOLIDS CONCENTRATE
TO PUG MILL
WATER 21,300 (MAX)
SOLIDS 2020
EXCESS TO RAW
WATER STORAGE

Figure 8-57. Lurgi gasification. Base case and alternate disposal and water recovery.

The capital cost for the same equipment, prorated from the Bechtel paper, is $3,990,000, providing a reasonably good check on the capital cost development calculation.

The capital cost of the base case evaporator was given previously as $4,000,000, thus the capital cost advantage of the alternate case is rather small. It should be noted that the Goslin evaporator capital costs are less than those indicated in the Bechtel paper (see "Costs of Water Treatment" section). Use of this reference would indicate a cost of $6,500,000. The operating cost of the alternate case, however, should be significantly lower, since the savings in low pressure steam usage is 48,675 lb/hr and credit for the steam saving should be only partly offset by the cost of the acid and caustic required to regenerate the demineralizer train.

Demonstration of the operability of reverse osmosis on waste sludges would be required, since the reject stream cannot contain precipitated solids that could clog the membranes.

Several alternates were considered because of concern about the possibility of high chlorides in the wastewater. The analyses employed in development of the Lurgi flowsheet of Figure 8-54 were from a different coal than that used by C. F. Braun in their designs on "western coal." Further, western coals are generally lower in chloride than eastern coals, particularly those from the Illinois basin. With Illinois coals the chloride content of the effluent water would probably be considerably higher than the 266 mg/l shown in the Lurgi flowsheet. The most stringent effluent standards, discussed earlier in the report, state that chloride shall not exceed 250 mg/l. Therefore, the 266 mg/l shown on the flowsheet would not meet this effluent discharge specification of 250 mg/l. Reverse osmosis is estimated to remove 90% of the chlorides and ion exchange could probably remove virtually all of it. Reverse osmosis would also lower the TDS level below the 1,000 mg/l effluent discharge specification and perhaps substantially reduce residual phenol.

More pertinent than effluent discharge specifications, since no discharge is contemplated, are the chloride, cyanide, phenol, sulfide, and TDS levels that must apply to reuse as cooling water makeup or makeup to high, medium or low pressure steam systems. These specifications were quoted earlier in this report and are repeated here in TABLE 8-63. It can be seen that the specifications are difficult to relate to specific compounds and do not easily lend themselves to illustration by tabulation. Such items as the Langelier and Ryznar indices relating to scaling have been omitted.

According to Betz Co. and others, the compounds shown in TABLE 8-64 cause difficulties in the cooling water system.

Foaming, usually caused by organic compounds in the water, can be troublesome in cooling tower operation, but can be controlled by use of anti-foaming agents. Several compounds, of which soda ash is an example, have been blamed for delignification of cooling tower wood. Cyanides are objectionable because of their highly toxic nature and should be removed or at least converted to less-toxic cyanates.

Many of the effects cited above can be reduced or eliminated by controlling pH and by the use of various additives. Many articles have been published on control of corrosion, scaling, foaming and fouling. Additives used in the past are being replaced or their use is being re-examined so that the rigid discharge controls for wastewater that have been introduced in recent years, or that are scheduled to be implemented, can be met.

TABLE 8-63. SPECIFICATIONS FOR MAKEUP WATER (1)

Contaminent	Cooling Tower Makeup	Boiler Feed Water 150 psig	Boiler Feed Water 600 psig	Boiler Feed Water 1,450 psig
	(Parts per million, unless specified)			
Chlorides	100-3,000 (2)	No spec., but undesirable		
Total Dissolved Solids (TDS)	2,500-3,000 (3)	60 ppb (3)	60 ppb	60 ppb
Suspended Solids	200-400 (3)			
Phenols	<1			
Ammonia	5			
Oils	<10			
Sulfide (as H_2S)	1-2			
Iron	0.5	0.10	0.025	0.01
Copper	0.08	0.05	0.02	0.01
Silica (4)	<150	150	30	< 2
Total Hardness (as $CaCO_3$)	Related to Indices	0.3	0.2	0.0
Sodium (5)	-	20 ppb	20 ppb	20 ppb
Total Alkalinity (as $CaCO_3$)	Carbonate <5 Bicarbonate 50-150	700	400	0

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(1) The cooling tower makeup specifications are unofficial recommendations from the paper "Reuse of Wastewater Effluent as Cooling Tower Make-up," Marvin Fleischman, U. of Louisville, WateReuse - 1975 (AIChE)

(2) Circulating water specification. Wide limits have been reported. 3,000 applies to stainless steel

(3) Circulating water specification

(4) < 20 ppb in saturated steam

(5) Specification in saturated steam, parts per billion

TABLE 8-64. PROBLEM COMPOUNDS IN COOLING WATER SYSTEMS

Compound	Corrosion	Scale Fouling	Biological Fouling	Reacts with Cooling Tower Chemicals
Ammonia	X		X	X
Oil			X	
Organics			X	
Phenols	X		X	
Suspended Solids		X		
Oxygen, CO	X		X	
Total Dissolved Solids		X		
Phosphorus		X	X	
Silica		X		
Sulfate		X	X	
Chlorides	X			
Free Mineral Acid	X			
Nitrate			X	
Sulfur Compounds	X	X		X
HCN	X			

The foregoing statements are paraphrases of information in the "Betz Handbook of Industrial Water conditioning", 7th Edition, 1976.

If reverse osmosis must be used to remove chlorides, the flows in the Lurgi flowsheet would change as shown in TABLE 8-65.

Cooling tower blowdown would be sent to the evaporator, as in the base case, together with the reverse osmosis reject.

The base case capital cost would be increased by the installation cost of reverse osmosis plus the incremental cost of the increased evaporator capacity or, as described earlier in the alternate to the Lurgi gasification base case, reverse osmosis plus demineralizer plus a small evaporator. The additional capital required for these two installations is roughly:

Reverse Osmosis	$1,258,000
Incremental Evaporator	1,093,000
Total	$2,351,000

Reverse Osmosis	$1,258,000
Reverse Osmosis + Demin. + Small Evaporator	640,000
	$1,898,000

The effects of chlorides and other corrosive materials in the cooling water system may be avoided through use of corrosion resistant materials in the cooling water system, including the heat exchangers using cooling water. Some conceptual designs (692) employ more than one cooling tower and cooling water system and confine wastewater makeup to the cooling tower that serves stainless steel exchangers. Such a solution for the corrosion problems is practical since many of the exchangers in gasification

TABLE 8-65. EFFECT OF REVERSE OSMOSIS IN THE LURGI FLOWSHEET

Stream	Water Flows Only, Lbs/Hr Base Case	Water Flows Only, Lbs/Hr With Reverse Osmosis
Leaving Biox Filter	1,295,200	1,295,200
To Cooling Tower	251,600	0
To Raw Water Storage	1,043,600	959,733
To Reverse Osmosis	0	335,467
From Reverse Osmosis		
Clean Water to Cooling Tower	-	251,600
Reject to Evaporator	-	83,867
Cooling Tower Flows		
Makeup from Reverse Osmosis	0	<251,600*
Makeup from Boiler Blowdown	280,900	280,900
Makeup from Biox	251,600	0
Evaporation and Drift	463,000	463,000
Blowdown (to Evaporator)	69,500	<69,500*
Total Evaporator Feed	283,700	<367,567

*The higher purity of the makeup would undoubtedly reduce the cooling tower blowdown requirement and thus reduce the makeup requirement, but actual reduction would have to be determined by operation or experimentation.

plants must be made of stainless steel or other corrosion resistant materials because of the corrosive nature of the streams on the process side of the exchangers.

Another approach might be stage-wise water condensation, as exemplified by the Braun HyGas design. The liquid condensing at the higher temperature would probably contain most of the inorganic materials, and this quantity of water could be treated at lower cost. It is our understanding that this approach is being investigated in the IGT HyGas pilot plant operation.

Other possible approaches include side-stream treatments on the cooling tower blowdown instead of on the makeup. These would be evaporation, reverse osmosis, demineralization, or combinations of the three as described for makeup treatment. Obviously, these would reduce makeup, and thus usage of the wastewater, by a maximum of 69,500 lbs/hr (zero blowdown). In order to prevent buildup of chlorides in the total water system, it will be necessary to remove at least as much as is introduced with the coal. Some Illinois basin coals contain as much as 0.5 percent by weight of chlorine. This represents 6,316 lbs/hr on the flowsheet basis which converts to 5,036 mg/l chloride in wastewater as a top maximum figure.

Another point worthy of mention in connection with high chlorides is that ammonia fixation (as NH_4Cl) would definitely occur and two-stage stripping with lime treatment between stages (see following section on gasification without p/o/t production) would be necessary.

Other inorganic compounds that may be found in coal gasification wastewater and that are not removed by lime precipitation, coagulation or flotation, are mainly boron compounds, sulfates, silicates, phosphates, and nitrates. Remarks and methods cited for chlorides would also apply to these compounds. Most are considered highly soluble but innocuous in a recycling water system unless they contribute to scaling or biological fouling. Should the water be discharged it must be treated by reverse osmosis and/or demineralization or evaporation to meet standards quoted earlier in the report.

Alternate methods for removal of organic compounds are numerous:

- There are many variations in biological treatment processes, such as trickling filters, rotating biological disc contactors, fluidized sand beds, use of high purity oxygen activated sludge, and others. It should be noted that Water Purification Associates (480) evaluated four variations: conventional, high purity oxygen activated sludge (HPOAS), trickling filter plus HPOAS and conventional plus nitrification and denitrification, and found the combination of trickling filter plus HPOAS to be the most cost effective.

- Anaerobic digestion in a first stage may be followed by one of the activated aerobic sludge processes. Phenol extraction by the Phenosolvan process could be eliminated with a capital cost saving of about $12 million.

  Envirotech has estimated that they could evaluate this option for $10,000 to $15,000 in existing equipment at their Salt Lake City laboratory.

- Use of powdered activated carbon in the biological oxidation system with wet oxidation regeneration (Zimpro-Dupont) was reported in a previous section to cost $8.5 to $11.5 million. This could be more than the $10 million estimated by Envirotech/Kellogg for the conventional air activated sludge system; however, it would undoubtedly be more efficient in removal of COD, BOD, cyanides,and thiocyanates than would the conventional system. The question is whether the cooling tower system could tolerate the greater amounts of these components present in the Envirotech conventional activated sludge effluent. There are a few precedents cited which indicate that the cooling tower is a good biological contactor itself and that the algae produced can be controlled. Some conceptual designs, e.g. El Paso, Burnham, are counting on this method of operation.

- Wet air oxidation may be considered as a substitute for first stage biological oxidation. Zimpro submitted an estimate of $12 to $16 million for this. Phenol extraction at a capital cost of about $12 million could be eliminated to offset this cost.

- The stream that is normally sent to biological oxidation may be sent instead to the central incinerator/boiler for the plant complex. This alternate appears to be viable technically, would save considerable investment (about $15 million, less incremental costs on the incinerator boiler) and would provide fuel to reduce coal consumption. Pullman Kellogg favors this option, but recommends testing, preceded by design estimates to check economic feasibility.

  Should the previous alternate prove to be too great a diluent for the incinerator/boiler feed, a separate catalytic oxidation of wastewater should be investigated.

- The academic case of treating for discharge, instead of reuse, would require addition of granular activated carbon treatment plus possibly ozone treatment to meet the very low organic effluent standards. In addition, the inorganic removal steps described earlier would be required (reverse osmosis and/or demineralization and/or evaporation).

Evaporation for Treatment of Condensates from P/O/T-Producing Gasification--

Vendors of evaporator equipment state that evaporation of wastewaters from gasification processes that produce p/o/t, after phenol extraction, stripping and flotation, appears to be feasible from an operational standpoint. Future development of control technology should include evaluation of evaporation of water containing organic compounds.

There are problems associated with this method of disposal:

- The quantity of organics in the evaporator discharge stream is estimated at about 1,000 lbs/hr in contrast to about 110,000 to 120,000 lbs/hr of total inorganic solid wastes that are normally discharged from the gasification plants for disposal, including ash/slag.
- Disposal of wastes containing organic compounds could cause problems in obtaining operating permits.
- Demonstration of disposal techniques for the organic-containing waste would probably be required on a long term basis on a demonstration plant scale.

Economics have been developed for the treatment scheme shown in Figure 8-54, the base case flowsheet for gasification processes producing p/o/t. The base case includes phenol extraction, stripping, and flotation followed by 2-stage biological oxidation and filtration of the clarified water for reuse as process water and cooling tower makeup. An evaporator, cold lime clarifier, and sodium softener are part of the overall water treatment scheme. Capital costs include $10 million for the biological oxidation system and $4 to 7 million for an evaporator processing 0.82 MGD of wastewater (Goslin estimated $4 to 4.4 million, while Bechtel indicated that the cost might be $6.5 to 7 million).

For the evaporator alternate case, the biological oxidation section is eliminated, the evaporator capacity is increased to 3.62 MGD and both the cold lime softener and the sodium softener are reduced. The evaporator capital cost, estimated by applying the 0.6 exponent to the ratio of capacities, becomes $9.7 to 10.7 million (averaging, say, $10.2 million) if Goslin figures are used, and $15.8 to 17.1 million if Bechtel figures are used. The question of multiple units versus single units has not been answered satisfactorily and requires further study: Bechtel describes the use of 3 evaporators for a total capacity of 2.33 MGD, while Goslin states that a single unit is feasible.

For the base case, the total capital for biological oxidation and the small evaporator ranges from $14 to 17 million. For the alternate case the capital cost of the evaporator ranges from $10.2 to 17.1 million. If Goslin is right, there is a possible capital cost saving with the alternate case of $3.8 million.

An additional capital cost saving in the alternate case may be realized, due to the reduction in cold lime softening and sodium softening requirements, of on the order of $650,000 for a total incremental cost advantage for the alternate case of $4,450,000.

Operating costs for the alternate case are probably higher than for the base case:

	Base Case	Alternate
Operating Costs without capital charges	$ 708,000	$1,456,000
Incremental capital at 15 %/yr	667,500	-
	$1,375,000	$1,456,000

Since none of the treatment alternatives have been demonstrated adequately, this evaporation alternate should be explored further. Evaporation equipment vendors, such as Goslin and Struthers Wells, have examined the specifications for the evaporator feed as supplied by Pullman Kellogg for the evaporator alternate case. The tentative conclusions are that evaporators can be made to function in this application without excessive foaming or corrosion but that testing should be carried out on actual samples.

There is some indication from unofficial telephone contacts that ElPaso Natural Gas has had some testing done. Most of the foaming appears to be confined to one of the six effects. Apparently some provision for scrubbing of carry-over organics is contemplated.

Bi-Gas/Koppers-Totzek Gasification Flowsheet, Figure 8-55

These processes produce no organic pollutants and a slag residue that is more leach-resistant than the ash residue produced by Lurgi gasification. Wastewater treating problems are thus much simplified. Modules shown by Braun on their flowsheet are:

- Sour water stripping to remove CO_2, H_2S,and ammonia
- Ammonia recovery (or removal) by Phosam-W
- Evaporation of inorganic sludges

Using the costing bases previously discussed, the cost for a Chevron WWT system processing 7.13 MGD is estimated as $18.1 million. This includes ammonia recovery, although the much smaller ammonia quantity, compared to Lurgi gasification, may justify only removal and not recovery for sale. Since ammonia interferes with the operation of the Claus process for sulfur recovery, it would have to be incinerated if it were not recovered. The decision for or against recovery would be specific for specific operating cases.

The evaporator cost (by Goslin) is estimated at $4.7 million.

Alternates/Critique of Base Case Gasification Producing No P/O/T

Ammonia removal is much more critical with gasification processes producing no p/o/t because biological oxidation is not included to utilize the residual ammonia. In the earlier discussion of stripping and ammonia recovery in the "Commercial Water Treating Methods" section the point was made that there is a good possibility that ammonia fixation will occur to prevent stripping to 20 to 50 mg/l residual ammonia unless pH is adjusted to 9.5 to 11 with lime or caustic. Therefore, Pullman Kellogg recommends a procedure published by Bethlehem Steel Corp. and used in their coke oven plants. Reference is "An Improved Process for the Removal of Ammonia from Coke Plant Weak Ammonia Liquor," E. M. Rudski, K. R. Burcaw and R. J. Horst, Iron and Steel Magazine, June, 1977. This reference reports excellent ammonia removal in a single steam stripper if this stripper is preceded by a clarifier/thickener fed from a pre-liming vessel where weak ammonia liquor is well mixed with sufficient 10 percent lime slurry to raise the pH to 11. Suspended solids from the clarifier/thickener were less than 50 mg/l while the underflow contained approximately 30 percent suspended solids. The underflow effectively removed tar or entrained coke. (The Bi-Gas process will probably entrain some char into the effluent water.) Subsequent stripping was free of deposits formerly encountered and less than 50 mg/l of residual ammonia was achieved with steam consumption of 0.13 kg/l WAL in contrast to the 0.30 kg/l WAL in the conventional stripping with lime addition to a "lime leg" communicating with the stripper. Pullman Kellogg believes that lime addition in this manner would also precipitate virtually all trace metals as well as other inorganic compounds which could contribute to problems in subsequent water reuse. Compositions of streams 2 and 3 on the bottom of the flowsheet reflect use of the above technique.

It would be necessary to adjust the pH of the stripped water to 7 to 8, and this could be done either with CO_2 produced in the gasification plant or with sulfuric acid.

Added cost of the clarifier/thickener may be offset by the smaller size of the stripper system in this alternate. We estimate the maximum added capital cost as $850,000 by adjusting similar costs furnished by Envirotech for clarifiers.

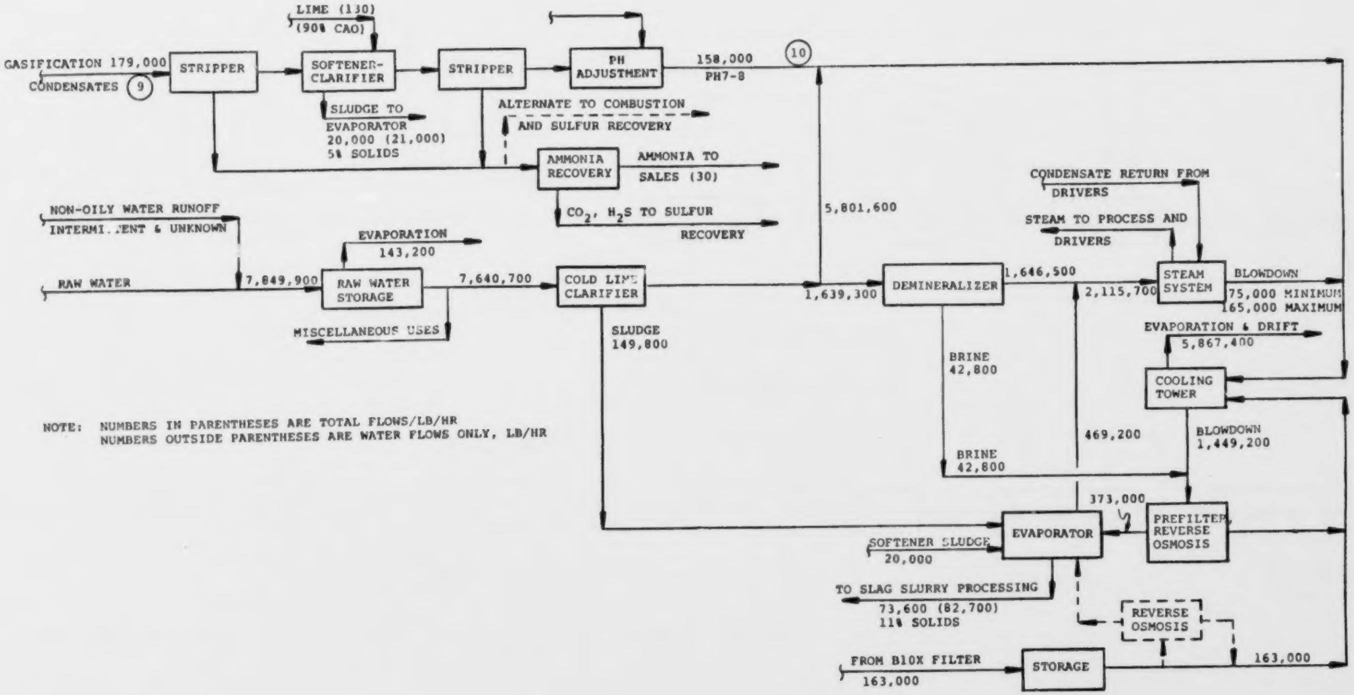
In order to use the water as cooling tower makeup or boiler feed water, combinations of reverse osmosis and/or demineralization and evaporation, as described for the Lurgi flowsheet, may or may not be necessary depending on the actual coal used and the concentrations of chlorine, boron, etc. that are present. Experimentation is necessary to establish whether ammonia, cyanides, and sulfides remaining might have to be further reduced by such methods as ozonation and/or activated carbon or alkaline chlorine treatment.

Liquefaction Prototype Flowsheet, Figure 8-56

As previously described, Figure 8-56 illustrates the method presented in the Parsons conceptual design of recycling stripped, combined wastewaters from the liquefaction reaction and the two high temperature gasifiers back to a waste heat boiler system feeding the process gasifier. Also as previously stated, we believe this should be tried on a demonstration plant to determine the fate of the inorganic compounds and whether their presence would lead to unacceptable scaling, clogging, or catalyst poisoning. It should be noted that this system contemplates a considerable discharge of wastewater to the river.

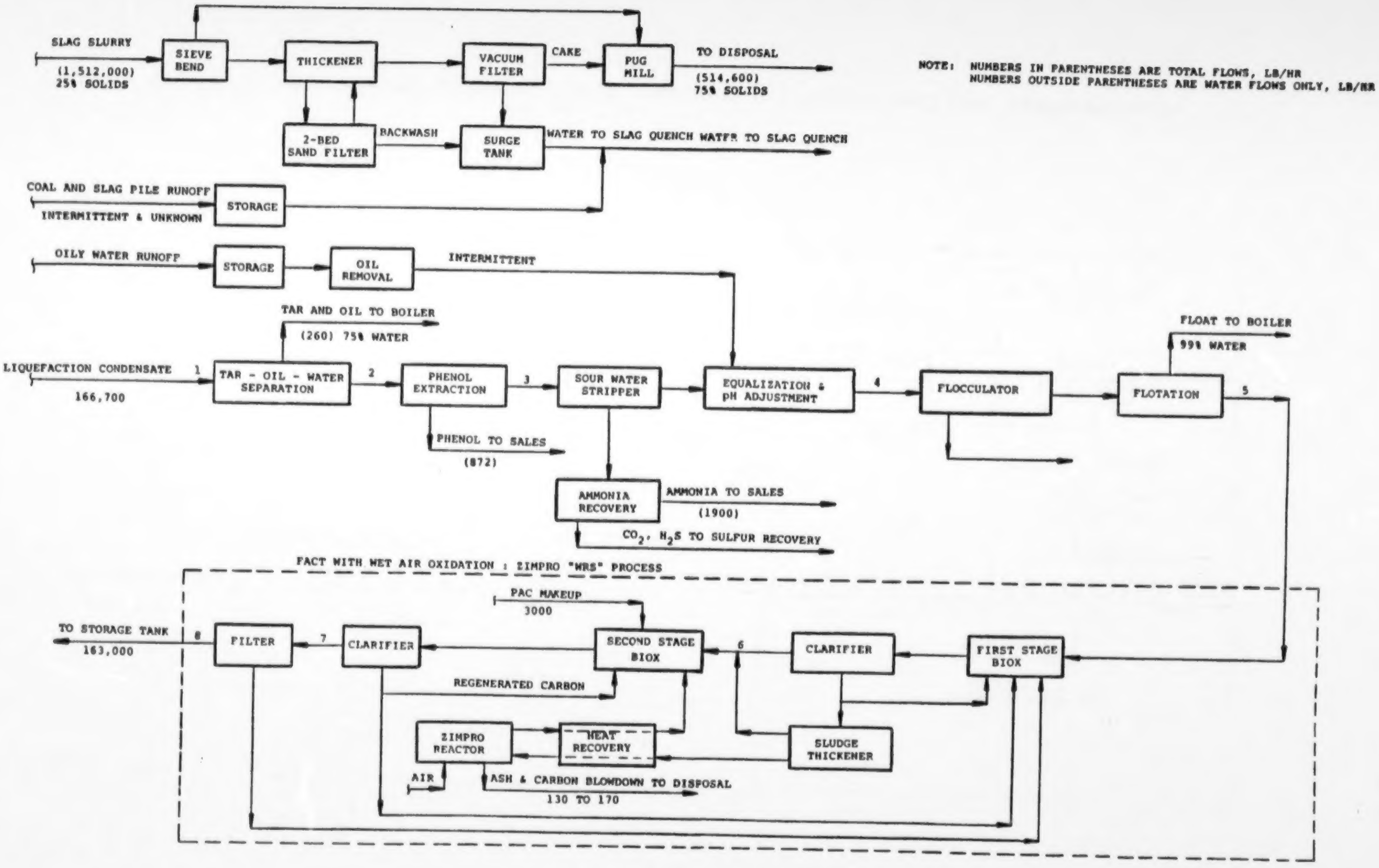
As an alternate to Figure 8-56, Pullman Kellogg presents a system more similar to the Braun gasification schemes; i.e., zero discharge. Figure 8-58 is an illustration of the Pullman Kellogg alternate. This scheme contains the following features:

- Separate, two-stage stripping of gasifier condensate with lime addition between stages to remove CO_2, H_2S, and ammonia to low levels. Stripped water is used for cooling tower makeup
- Oil separation on the liquefaction wastewater
- Phenol extraction of "oil-free" liquefaction wastewater
- Single stage stripping and ammonia recovery of extracted water using either Phosam-W or Chevron WWT proprietary processes. Although adjustment of pH may not be necessary prior to stripping, caustic addition is provided
- Equalization and pH adjustment to 7 to 8
- Flocculation
- Flotation
- Two-stage biological oxidation with powdered carbon addition


GASIFICATION 179,000
CONDENSATES ⑨
STRIPPER
LIME (130)
(90% CAO)
SOFTENER-CLARIFIER
SLUDGE TO EVAPORATOR 20,000 (21,000) 5% SOLIDS
STRIPPER
PH ADJUSTMENT
158,000
PH7-8
⑩
ALTERNATE TO COMBUSTION AND SULFUR RECOVERY
AMMONIA RECOVERY
AMMONIA TO SALES (30)
CO2, H2S TO SULFUR RECOVERY
NON-OILY WATER RUNOFF
INTERMITTENT & UNKNOWN
RAW WATER
7,849,900
RAW WATER STORAGE
EVAPORATION 143,200
MISCELLANEOUS USES
7,640,700
COLD LIME CLARIFIER
SLUDGE 149,800
5,801,600
1,639,300
DEMINERALIZER
1,646,500
BRINE 42,800
BRINE 42,800
2,115,700
469,200
CONDENSATE RETURN FROM DRIVERS
STEAM TO PROCESS AND DRIVERS
STEAM SYSTEM
BLOWDOWN
75,000 MINIMUM
165,000 MAXIMUM
EVAPORATION & DRIFT 5,867,400
COOLING TOWER
BLOWDOWN 1,449,200
373,000
PREFILTER, REVERSE OSMOSIS
EVAPORATOR
SOFTENER SLUDGE 20,000
TO SLAG SLURRY PROCESSING 73,600 (82,700) 11% SOLIDS
REVERSE OSMOSIS
FROM BIOX FILTER 163,000
STORAGE
163,000

NOTE: NUMBERS IN PARENTHESES ARE TOTAL FLOWS/LB/HR
NUMBERS OUTSIDE PARENTHESES ARE WATER FLOWS ONLY, LB/HR

Figure 8-58a.- Integrated scheme for treatment of liquefaction wastewaters.


SLAG SLURRY
(1,512,000)
25% SOLIDS
SIEVE BEND
THICKENER
VACUUM FILTER
CAKE
PUG MILL
TO DISPOSAL
(514,600)
75% SOLIDS
2-BED SAND FILTER
BACKWASH
SURGE TANK
WATER TO SLAG QUENCH WATER TO SLAG QUENCH
COAL AND SLAG PILE RUNOFF
INTERMITTENT & UNKNOWN
STORAGE
OILY WATER RUNOFF
STORAGE
OIL REMOVAL
INTERMITTENT
TAR AND OIL TO BOILER
(260) 75% WATER
LIQUEFACTION CONDENSATE
166,700
1
TAR - OIL - WATER SEPARATION
2
PHENOL EXTRACTION
3
SOUR WATER STRIPPER
EQUALIZATION & pH ADJUSTMENT
4
FLOCCULATOR
FLOTATION
5
FLOAT TO BOILER
99% WATER
PHENOL TO SALES
(872)
AMMONIA RECOVERY
AMMONIA TO SALES
(1900)
CO2, H2S TO SULFUR RECOVERY
FACT WITH WET AIR OXIDATION : ZIMPRO "WRS" PROCESS
PAC MAKEUP
3000
TO STORAGE TANK
163,000
8
FILTER
7
CLARIFIER
SECOND STAGE BIOX
6
CLARIFIER
FIRST STAGE BIOX
REGENERATED CARBON
ZIMPRO REACTOR
HEAT RECOVERY
SLUDGE THICKENER
AIR
ASH & CARBON BLOWDOWN TO DISPOSAL
130 TO 170

NOTE: NUMBERS IN PARENTHESES ARE TOTAL FLOWS, LB/HR
NUMBERS OUTSIDE PARENTHESES ARE WATER FLOWS ONLY, LB/HR

Figure 8-58b.- Integrated scheme for treatment of liquefaction wastewaters.

	Streams in Figure 8-58									
	1	2	3	4	5	6	7	8	9	10
	(mg/l unless otherwise noted)									
BOD	52700	52700	42700	9100	9100	455	45	40	~0	~0
COD	88600	88600	73400	14200	14100	3700	1000	950	128	52
TDS	5300	5300	5300							
TSS	2	2	2	2	2	20	20	5	5081	5
Phenol	6800	6800	410	410	410	20	< 1	< 1	0.011	0.011
Cyanide	10	10	10	7	5	0.5	< 0.05	< 0.05	13	6
Thiocyanate	350	350	350	350	350	100	35	35		
Ammonia	14400	14400	14400	45	45	25	1	1	184	20
Sulfide	29300	29300	29300	10	10	2	0.06	0.06	7	7
Chloride	100	100	100	100	100	100	100	100	100	100
Oil	608	220	50	50	25	10	5	1	~0	~0
Hardness*	<80	<80	<80	<80	<80	<80	<80	<80	630	80
Alkalinity*	80000	80000	80000	700	700	700	700	700	650	10
pH	9.5	9.5	9.5	7.5	7.5	7.5	7.5	7.5	8.5	7.0
Water flow, lb/hr	166700	166400	166400	166400	166000	76300	163000	163000	179000	178000

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*As $CaCO_3$

Figure 8-58(b). Stream compositions for integrated scheme for treatment of liquefaction wastewaters (Part 3 of 3).

and regeneration of circulating stream by wet oxidation (Zimpro and DuPont PACT). Sludge handling as indicated with concentrated sludge to the central incinerator/boiler

- Multi-media filtration of biological oxidation effluent prior to use as cooling tower makeup
- Installation of reverse osmosis plus evaporation of its reject and the inorganic sludges from the cold lime clarifier and demineralize. Boiler blowdown is used as cooling tower makeup to reduce fresh water usage

"Zero Discharge" is thus achieved except for the combined wet (ca. 25 percent water) slag-and-inorganic residue stream and cooling tower evaporation and drift.

Using the costs in the sections on "Costs of Water Treatment" and "Budget Cost Estimates Received From Licensors and Vendors," our best estimate of the capital cost of the system shown in Figure 8-58 is developed in TABLE 8-66.

Parsons (804) indicated the following costs in their report:

Raw Water Treating	$15.48 MM
Sour Water Treating	5.53
Effluent Water Treating	5.21
Total Water Treating	$26.22 MM

Only the last two items, total $10.74, might be compared with the $33.506 for our treating system; however, we eliminate most of the Parsons demineralizer with evaporation. A more direct approach is to evaluate those items eliminated from the Parsons scheme by Pullman Kellogg:

TABLE 8-66. CAPITAL COST OF ZERO DISCHARGE SYSTEM FOR LIQUEFACTION WASTEWATER TREATMENT

Item	Capacity, MGD	Capital-$MM	Source/Reference
Gasifier Condensate			
Stripping/NH_3 recovery	0.49	3.73	Chevron Research
Lime clarifier	0.49	0.34	Kellogg/Envirotech
Liquefaction Condensate			
API Separator	0.49	0.16	SRI (804)
Phenol Extraction	0.49	5.07	American Lurgi
Stripping/NH_3 recovery	0.49	3.73	Chevron Reserach
Equalization, pumping station	0.49	0.16	Pullman Kellogg
Flocculation, Flotation	0.49	0.436	Kellogg/Envirotech
Two-Stage Biological Oxidation with PACT (Filter Included)	0.49	9.00*	
Reverse Osmosis (Including pre-filter)	4.59	5.035	L*A Water
Evaporator	4.146	5.915	Goslin/Envirotech
Total Capital Cost		33.506	

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*Zimpro says lower hydraulic flow with the same BOD load has very little effect on cost.

Reduction in Demineralizer	$1.47 MM
Eliminate 4.34 MGD clarifier	0.69
Eliminate sludge filter (0.44 MGD)	0.11
Reduce size of raw water treating ca. 20%	2.66
Total savings	$4.93 MM

Adding the pluses and minuses:

Total Parsons Treating Cost	$26.22 MM
Units eliminated	- 4.93
	$21.29 MM
Deduct Parsons' sour water stripping	- 5.53
Parsons system remaining	$15.76 MM
Add Pullman Kellogg system	33.51
Total treating system	$49.27 MM
Total Parsons system	26.22
Net added by Pullman Kellogg	$23.05 MM

Some additional cost may be eliminated from the Parsons system by reducing the waste heat boiler injection system costs, but this cost cannot be estimated without much more detail than is available at this time. Consequently, what has been presented in effect is the maximum (Pullman Kellogg) and minimum (Parsons) capital costs in handling liquid effluents from a coal liquefaction plant: $49.27 MM vs. $26.22 MM.

It is our recommendation that any demonstration plant built should be the conservative Pullman Kellogg system, but should include design features to permit testing of the simpler Parsons system of injecting stripped combined wastewater into the waste heat boiler system that feeds the process gasifier.

The same alternate treating systems for inorganic and organic matter enumerated for the Lurgi system would apply also to the liquefaction wastewater, so these will not be reiterated in this segment of the report.

Operating Costs of Integrated Systems

Partial operating costs for the integrated systems can be developed by use of the information tabulated in the segment on operating costs for the separate treatment methods. These operating costs are shown in TABLE 8-67 for the recommended systems for p/o/t gasification, gasification with no p/o/t, and liquefaction.

The treatment units enumerated in TABLE 8-67 are the major systems contributing to operating costs. Processes such as API separators, pH adjustment, flocculation and flotation would add relatively minor operating costs to those shown. Such operating costs as are included in these operations would probably be mainly chemical costs rather than utility or amortization costs. Amortization costs for these operations can be computed at 15 percent of capital per year, to develop the added costs as follows:

Gasification with p/o/t	$827/day
Gasification with no p/o/t	None
Liquefaction	$311/day

Inclusion of chemical quantities for pH adjustment, lime, alum, and polyelectrolyte would be speculative without experimentation, since the process requirements are not well enough known.

TABLE 8-67. OPERATING COSTS OF INTEGRATED SYSTEMS

Operating Costs in $/day (1)

Treatment Unit	Gasification with p/o/t	Gasification no p/o/t	Liquefaction
Phenol Extraction (using avg. values)	$ 7,690	None	$ 2,318
Chevron Stripping & Ammonia Recovery	13,982	$25,334	3,253 (2)
			3,253 (2)
PACT Biological Oxidation (Zimpro-DuPont)	6,605	None	2,276
Reverse Osmosis (Dow)	-	-	2,071
Evaporation (Goslin)	2,525	2,913	3,669
Partial Total	$30,802	$28,247	$16,840

---

(1) Includes capital-related items at 15 percent of capital per year

(2) Caustic or lime costs not included

Raw water treating systems, such as lime softening, zeolite softening,and demineralization, are not included in this analysis, although they were included on the flow sheets for illustrative purposes and to show quantities, because of the extremely great variation in raw water analyses encountered in the areas that are favorable for location of conversion plants.

EFFICIENCY OF WASTEWATER TREATMENT SCHEMES

The water treating schemes outlined in the preceding section assume "zero discharge" to receiving waters. A discussion of whether or not effluent waters can meet the most stringent of the standards for discharge to receiving waters is therefore academic. The criteria for zero discharge are:

- Wet sludges and ash must be properly impounded so there will be no leaching into receiving waters via rainfall.
- Available water treating methods must be efficient enough to permit recycle of wastewater for cooling tower or boiler feedwater use without creating unacceptable corrosion, foaming, algae formation, heat exchange surface fouling, or adverse reactions with cooling tower or boiler feedwater chemicals.

Based on available data on the compositions and quantities of conversion process effluents and on data and information concerning water treatment technology, we believe that both criteria can be met. However, proof that the criteria can be met requires demonstration by applying the techniques suggested and evaluating the results over an extended period of time.

Lurgi plants in the U.S. are planned by Wesco, El Paso, ANG, and others, and it is assumed that final designs for such plants will include pollution control technology based on best engineering judgement of the data available at the time of design. Operation of the plants will prove or refute the design judgements. One or more of these projects appear to offer the best near-term opportunity for the study of water reuse technology; however, DOE demonstration plants may precede the commercial Lurgi plants into operation. In any case, intensive study of minimum treatment of water necessary to permit recycle to cooling tower or boiler systems without unacceptable corrosion, foaming, algae accumulation and fouling, scale accumulation and fouling, etc. should be carried out on the first plants large enough for the results to be meaningful.

Once the large plants are in operation, water treating licensors and equipment vendors should be issued samples or granted access to the cooling tower and boiler systems to study the requirements for both total recycle and further treatment to meet effluent quality standards.

Developing Technology

A number of newer water treating processes have been disclosed in the open literature or are being touted in advertisements by licensors and vendors. Testing and evaluation would be necessary to determine if any of these offer any real advantages over existing technology in coal conversion processes. Aspects of this newer technology could be:

- Energy savings
- Reduction in capital and operating costs
- Additives to prevent fouling, scaling, foaming, etc.
- More efficient removal of contaminants
- Removal of any special contaminants introduced by a particular coal, a particular raw water or a prior treating process
- Elimination of operating problems

- Reduction of disposal costs

Promising developing technology in water treating which we have noted, and the source of the information, includes the following:

- Application of fluidized-bed technology to biological oxidation.
  We found references to this technique of extending surface for biological activity. Brochures and a letter from Ecolotrol, Inc. claims a developed system with far less space requirement than conventional biological oxidation. The brochure states that pilot plant rental units are available. A technical paper describing a unit in operation in Nassau County, N.Y. appeared in the May 1977 issue of the Journal of Water Pollution Control Federation. Oak Ridge National Laboratory published results of testing ("Biodegradation of Phenolic Waste Liquors in Stirred Tank, Columnar, and Fluidized-Bed Bio-Reactors" CONF-761109-4, Holladay, et al., AIChE meeting, Chicago, Nov.28 - Dec. 2, 1976. Kellogg reference 610). The fluidized bed exhibited highest degradation rate and phenol effluent of 0.5-1.0 mg/l, but retention time was not high enough for good thiocyanate conversion.

- Oil fluidized evaporation.
  This is a technique for removing water from waste streams, leaving the solids behind suspended in oil. The oil is separated from the solids in a centrifuge and reused. The solids are encapsulated, incinerated, or otherwise prepared for ultimate disposal. This technique, called the Carver-Greenfield process, claims the advantages of eliminating fouling of evaporator heat

exchangers and of production of dry solids. The process is marketed by Dehydro-Tech Corp., East Hanover, N.J., from whom brochures and correspondence have been received. Both mobile and stationary pilot facilities are available.

- The Sulfex process for removing heavy metals from waste streams.

  Developed by the Permutit Company, Paramus, N.J. (Chemical Engineering, May 9, 1977), the process is claimed to out-perform conventional hydroxide precipitation while avoiding H_2S evolution. A prototype was to begin operation by end 1977. It should be mentioned that several vendors we talked to mentioned use of sodium sulfide as a scavenger in water treating systems as a conventional technique. Most inorganic sulfides apparently can be precipitated from aqueous solutions.

- Ultraviolet irradiation.

  Several references, among these Chemical Engineering of Aug. 1, 1977, have been noted regarding ultraviolet irradiation. Irradiation of ozone-enriched secondary treatment wastewater converted all organics to CO_2 and H_2O, heavy metals precipitated as either oxides or metals and PCB's and viruses were destroyed.

- Sludge volume reduction.

  A number of vendors offer lime sludge recovery systems, e.g., Dorr-Oliver's Fluo-Solids lime sludge recovery, which could be of interest to determine the cost effectiveness of the process. Other techniques for reducing sludge volume include Permutit's "Spiractor," which works by "catalytic precipitation" and produces hard pellets of sludges.

- Freezing.

  Freezing has been mentioned as a promising technique, but few data are available. Fluor (PB-273 318/6WP "Comparative Economics of Freezing Process as Brine Concentrators" Shroeder, et al.,June 1977) prepared cost estimates to compare with vertical tube vapor compression evaporators using $CaSO_4$ seeding to prevent scale formation. The method appears feasible but requires development.

- Treatment of purge streams that are toxic to bacteria.

  Reagents purged from sulfur removal systems have been reported to kill the working bacteria in biological systems (SRC, personal communication, and monthly report from Ft. Lewis, Wash. pilot plant) and therefore should not be allowed to reach biological oxidation feeds. NCE (Nittetu Chemical Engineering, Ltd.) sells a thermal incinerator for these reagents which operates in a reducing atmosphere. Purge liquors from redox systems, such as Takahax and Stretford, can be treated and recycled back to the process.

- Catalytic oxidation.

  Catalytic oxidation of organic-laden wastewater may prove to be viable and should be investigated further. A good reference is a paper from the Proceedings of the 29th Industrial Wastewater Conference at Purdue University, May 7-9, 1974: "Aqueous Phase Catalytic Oxidation as a Wastewater Treatment Technique," Kotzer, et al. Copper oxide catalyst was employed at oxygen partial pressure of 6.8 atmospheres at 100 to 300° C, using air with 10 percent excess oxygen. Reported results of treatment of coke plant wastes containing 4000 mg/l organics showed that 99 percent or more of the organics were destroyed. The cost was estimated at 75¢/1,000 gal on an 800,000 gal/day plant. An expander was employed on the exit gases to run the air compressor.

- Removal of oil from emulsified mixtures.
  From the same conference as the item above, an interesting paper on removal of oil from emulsified mixtures was presented. This could be important in liquefaction and gasification processes producing p/o/t. Reference was entitled "Removal of Oil from Dilute Aqueous Emulsions by Auto-Coacervation and Coalescence in Carbon-Metal Granular Beds," by Brown (Nalco) and Ghosh (U. of Maine). The technique was said to be especially suitable for removing "oil haze."

  We are concerned that the API separator shown on the prototype flowsheets is not sufficiently sealed to prevent air leakage in or noxious gases out. A closed coalescer bed such as described above might prove to be the best solution to this concern. It should be mentioned that American Lurgi has a solid bed guard chamber to reduce oil entry to Phenosolvan and Chevron WWT says any air contact would likely interfere with stripping, especially of sulfides.

- Removal of chlorine.
  Chlorine is expected to be a problem with eastern coals. No data are available on the composition of the chlorine-bearing materials. Should the chlorine combine with organics, an article of interest is "Choosing a Process for Chloride Removal" by M.F. Nathan (Crawford and Russell) (Chemical Engineering, Jan. 30, 1978). Light organic chlorides can be removed by stripping. Aromatic chlorides are best removed by carbon adsorption or polymeric adsorption. Other methods discussed are biological

oxidation and extraction. Data are presented.

- The Lindman precipitator.
  Precipitator, Inc., Santa Fe Springs, Col. makes interesting claims for the Lindman precipitator (Chemical Engineering, Jan. 2, 1978). This is a physical and chemical wastewater treatment that uses sulfur dioxide, iron and lime in a continuous flow process. On a diluted primary digestor sludge reductions were: TDS, 80.8 percent, SS, 99.4 percent, BOD, 73.9 percent and oil and grease, 93.1 percent. Capital equipment cost is about $1/gal/day for units larger than 250,000 gal/day. Operating cost is 20¢/1,000 gal excluding labor. Chemical usage is 2 pounds each of lime and SO_2 and 1 pound of iron per 1,000 gallons of wastewater.

- Wet Oxidation of low grade fuels and wastes.
  Zimpro, Inc. sent us a technical paper on "Wet Oxidation Boiler-Incinerator" by L.A. Pradt. Mass and energy balances, performance, and costs are described for a system of steam generation using low grade fuels and wastes. Advantages include no SO_x, NO_x, metal, or halide vapors, negligible particulates in stack gases and ability to use wet fuels, liquors, and sludges. Operating costs for 100,000 lb/hr of steam are $4.34/1,000 lbs. With sludge disposal credit, the cost drops to $2.70/1,000 lbs. Installed cost is estimated at $7,400,000. Base fuel is coal.

- Sludge pyrolysis.
  Another technique which may have application, or possibly show some advantages, is sludge pyrolysis (see ENR, Oct. 6, 1977, "Pure Water"). Nichols Engineering and Research Corp. offers systems for processing municipal sludge.

- Liquid membranes.

  An example of a process under development and only announced in technical literature is "Wastewater Treatment by Liquid Membrane Process," Kitigawa and Nishikawa (Takuma Co., Osaka, Japan) and Frankenfeld and Lia (Exxon Research and Engineering), Environmental Science and Technology, June 1977. Lab and pilot plant studies show liquid membranes "capable of reducing levels of NH_4^+, hexavalent chromium, copper, mercury, and cadmium from several hundred ppm to less than 1 ppm."

- Mercury removal.

  Another reference describing Japanese know-how is "Mercury Clean-up Routes-II," by Nicholas Iammartino (Chemical Engineering, Feb. 3, 1975). The Japanese process referred to in the article is termed "re-elixirization" and involves mixing wastewater with a divalent ferrous salt, neutralizing with alkali, and oxidizing with air. Magnetic separation then removes insoluble ferromagnetic ferrite for disposal or special uses. In two units mercury was reduced from 6.0 and 7.5 mg/l to 0.005 and 0.001 mg/l. Arsenic, chromium, lead, cadmium, zinc, copper, and magnesium were also reduced to low residual values. Some coals contain mercury and gasification process data have shown mercury to volatilize and condense into the product water. The re-elixirization process, or the other two processes mentioned in this article (FMC and Georgia-Pacific) could conceivably be required should mercury prove to be a problem in the more conventional processes being proposed.

- New bacteria strains.

  "Super Bacteria" bred by Polybac Corp. of New York, N.Y. are offered commercially. They appear to be especially good for reducing ammonia, cyanides, oil, phenols, detergents and toxic organics to levels much lower than conventional bacteria. Reference is "Plant Engineering," June 23, 1977. Costs are said to be 1.5 cents/1,000 gallons. Application as a "clean-up" biological stage or for startup, high-load periods, or to aid in shock recovery, should be investigated.

Water recovery in reverse osmosis systems may be enhanced by a method that is described in "Significantly Increased Water Recovery From Cooling Tower Blowdown Using Reverse Osmosis" by L. J. Kosarek (El Paso Environmental Systems, Inc.) (884). The paper was presented at the Atlanta AIChE meeting Feb. 26-March 1, 1978. Experimentation is recommended by DOE on pilot plant or demonstration plant cooling towers. Briefly, cooling tower blowdown and plant effluent are combined in a holding pond. Water is pumped from the pond to a blend tank where temperature, pH and "anti-scalant levels" are controlled. One such antiscalant mentioned is a polyphosphate. Blend tank effluent is filtered and sent to reverse osmosis. A spiral wound reverse osmosis membrane is recommended. This procedure is said to increase water recovery over more conventional methods of reverse osmosis.

Even further recovery is said to be obtained if "a chemical feed" which inhibits and counteracts the anti-scalant in the brine is added. A clarifier is necessary to remove calcium sulfate precipitate and supernatant is returned to the blend tank. The precipitate is also said to remove dissolved silica and magnesium sulfate by coprecipitation or adsorption on calcium sulfate crystals. A "small bleed stream" purge is necessary, which can be sent to a solar pond or evaporator. Examples are cited where the described system improved reverse osmosis recovery from a usual 75 percent to 90 percent, from a usual 34 percent to 85 percent and even from a usual 10 percent to 90 percent on a utility plant effluent and blowdown. The paper confirms our previous statements that reverse osmosis is considerably cheaper than evaporation:

	Reverse Osmosis	Single-effect Evaporator
Capacity, gpd	144,000	144,000
Investment	$200,000	$370,000
Operation (labor, chemicals, maintenance, overhead)	$ 1.95/M gals	$ 1.95/M gals
Depreciation & Taxes (11%)	$ 0.79/M gals	$ 0.84/M gals
Electricity (2¢/KWH)	$ 0.10/M gals	$ 0.21/M gals
Cooling Water (4¢/Mgals)	-	$ 0.54/M gals
Total except steam	$ 2.84	$ 3.54
Steam ($3.00/M lbs)*	-	$ 13.63
Water Recovery	80%	95%
Cost, per/1000 gals recovered	$ 2.84	$ 14.40

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*This could be considerably reduced by multi-effect evaporation. Charging for use of low pressure steam @$3.00/1000 lbs may not be fair when backpressure steam from turbines is available.

These are but a few of the proprietary and developing technologies that have come to our attention. Whether they will be needed for special cases or will supplant older technologies must await the actual testing of conventional means as recommended earlier in this section. The newer techniques may then be applied to any remaining shortcomings of the older technologies.

NEED FOR LABORATORY DATA ON AVAILABLE SCHEMES

Although Pullman Kellogg has presented estimates based on best engineering judgement on the efficiency of various water treating methods, it is obvious that virtually all of the schemes presented must be verified. Actual coal conversion process wastewaters should be supplied to licensors and vendors for testing in their own laboratories or rental treating equipment supplied by the vendors should be operated directly in the pilot plants.

Potential problems that should be resolved are outlined by treatment method category in the following discussion.

Oil Separation

Laboratory investigations are needed to more clearly determine whether there are emulsion problems and, if problems are apparent, the best means to break the emulsions. Equipment vendors could be of great help in this respect.

Reports of pilot plant operations have not included in clearly usable form the concentrations of fly ash, char, unconverted coal fines,or other insoluble solids that may be present in the sour water. Information and data are needed from which any effects of the solids on oil separation may be ascertained and the best means to deal with the solids may be determined.

Oil separation means other than API separators ahead of the stripping operation would have to be enclosed to prevent H_2S and ammonia evolution. We believe it is desirable to strip as soon as possible unless enclosed processes such as phenol extraction are employed.

Stripping

Analyses both before and after stripping have not been located so far. The Carnegie Mellon/AGA project should publish some analyses soon. We believe single stage stripping with steam will drive CO_2 and H_2S to low levels (5 to 10 ppm) and ammonia at least to the level that is required for biological oxidation (100 to 300 ppm). This statement requires the confirmation of laboratory work on actual waters. In addition, the amount of ammonia actually required for biological oxidation must be established for use as a guide for the stripping investigations. Our theory is that the biological pond need have no more ammonia than can be stoichiometrically used by phenols and other easily biodegradable compounds in the first stage. Cyanates apparently do not begin to degrade until these compounds are gone and when cyanates degrade they produce ammonia. Residual ammonia from the final biological stage will be hard to control unless long sludge age is used, probably with powdered carbon.

The problem of obtaining low ammonia residuals when biological oxidation is not used can be solved, we believe, by two-stage stripping with lime clarification between the stages. Lime addition to pH 9.5 to 11 will be beneficial in many ways: it precipitates tars, suspended solids and trace metals as well as freeing "fixed" ammonia from ammonium salts of acids, such as ammonium chloride. Early simulation of this method on process condensate derived from coal conversion processes operating at high temperature and containing no p/o/t is recommended.

Flotation

Following second-stage stripping, recarbonation with carbon dioxide and addition of sulfuric acid should be compared to determine the best method of pH adjustment before flotation and biological oxidation. Flotation and biological oxidation may each require a different pH: flotation is said to be favored by somewhat acid pH. The proper additives to obtain best oil removal should be established by laboratory test with vendor participation.

Biological Oxidation

This process definitely requires piloting to determine the actual residuals of ammonia, cyanide, thiocyanate,and other compounds. High surface area powdered activated carbon should be evaluated to establish the improvement in the lowering of residuals at different levels of sludge age and carbon content of sludge.

Activated Carbon

Granular carbon beds following biological effluent filtration should be piloted or tested in the laboratory to clearly establish the residuals of contaminants that may be reached by this method in comparison to the use of powdered activated carbon.

Regeneration of powdered activated carbon in biological sludge by wet oxidation is another variation which merits investigation.

Reverse Osmosis

Inorganic and organic removal is possible by reverse osmosis. Removal of organic residuals such as soluble oil, phenols, cyanides and cyanates should be investigated as well as removal of inorganics such as chlorides, boron, ammonia, and others.

Liquefaction Wastewater Disposal

Both Parsons and COGAS Development Co. (see U.S. patent 3,966,633) have published design conceptions in which stripped wastewater from liquefaction is injected into a high temperature gasifier, which produces slag and not phenolics, in order to destroy the phenolic and other organic impurities. This appears to be an excellent means of disposal of the wastewater and therefore provisions should be made to test the scheme on an integrated liquefaction pilot plant or demonstration plant.

It would be interesting to determine the fate of the inorganic components of the wastewater. Do they build up and cause trouble with catalyst plugging or poisoning? Do they they cause scaling or plugging of exchangers or pipes? Do they eventually work themselves out in the slag without causing any of these problems?

It would also be interesting to have the real cost breakdown of the very large waste heat boiler system that Parsons employs to vaporize the wastewater into the gasifier.

Cooling Tower Operation

We have consulted various cooling tower experts in order to determine the effects on tower operation of residual amounts of ammonia, cyanides, cyanates, sulfides,and various inorganic or organic compounds that may not be removed from the wastewater that is used for tower makeup. No really satisfactory answers can be obtained without fairly long-term experimentation, probably on the demonstration plant scale. However, supplying cooling tower vendors and specialists with actual samples from pilot plant operation would be helpful to them. Certain tests could be made in their laboratories and certain parts of the problem are amenable to calculation such that recommendations on programs for prevention of corrosion and scaling could be made for the demonstration plant.

Betz Company personnel stated that they would prefer to be called in during the design phase of the cooling system. Certain techniques in exchanger design avoid later operational troubles. For example, valve installations to permit flow reversal and "air bumping", both without taking the exchanger off the line, have been found to be practical in dislodging scale accumulations before the scale severely interferes with heat transfer.

If foaming appears to be a problem, its severity could be determined in small-scale apparatus and antifoam agents could be evaluated. Similarly, prediction of biological (algae) fouling could be facilitated and additives recommended. Several firms offer consulting and design services for nominal fees and arrangements for DOE funds and for samples or plant access should be made.

Most of the same firms also have boiler feedwater preparation know-how. However, these methods are generally better known (ion exchange, reverse osmosis, evaporation, etc.) and they are known to remove practically all contaminants to tolerable levels if properly applied and, where necessary, preceded by appropriate pretreatment. The large body of knowledge from power plant practice is applicable to boiler feedwater preparation in coal conversion plants.

DOE Treating Programs

A number of instances have been cited previously of DOE coal conversion pilot plant water treating facilities or testing done or in progress by DOE contractors or subcontractors. These are summarized in the following discussion.

Solvent Refined Coal--

The 50 TPD pilot plant at Fort Lewis, Washington has the following train of equipment: surge reservoir, clarifier, dissolved air flotation, biological oxidation, sand filter, and charcoal filter. Unfortunately, this unit has been operated with large dilution streams, principally 216,000 gpd of belt cooling water which could be recycled back to process with minimum treatment and kept out of the treating train. It would be most interesting, once dilution has been reduced, to sample each step of the treatment scheme for influent and effluent analysis and analysis of sludges produced. To our knowledge this has not been done, or at least no reports have been published. One monthly report revealed that purge from the Stretford unit poisoned the bacteria in the biological oxidation unit and this purge is now sent to separate disposal. Other inorganic sludges are likewise sent to separate disposal:

- Thurston County Land Fill gets off-specification coal and SRC, waste sand from filters, waste charcoal, and asphalt coated rocks.
- Land Farm (Ft. Lewis) receives alum sludge, clarifier, and DAF skimmings, and excess biological oxidation sludge.
- Western Processing, Kent, Wash., a hazardous waste disposal site, receives ash, Stretford purge and "tank bottoms."

Sanitary waste and refuse are handled by the Fort Lewis municipal systems.

We feel that a good opportunity may exist here for study of ways of consolidating these wastes and determination of the best solid waste management system to isolate the solid residues from ground water. Various solid stabilization systems, such as Chem-Fix, IU Conversion and the like, could be evaluated as well as evaporation, filtration and reverse osmosis to concentrate the solids.

High Btu Gasification--

DOE sponsors, and partially funds with AGA, the Carnegie-Mellon

University project that monitors the HyGas, Bi-Gas, CO_2 Acceptor, Synthane and Grand Forks slagging gasifier pilot plants. This team has conducted some treating experiments but reports have not been published to date. No doubt, this knowledgeable group can solve in the laboratory some of the problems that have been cited. Pullman Kellogg submitted several ideas to the Carnegie-Mellon team, have met with them, talked by telephone and corresponded with them to discuss possible laboratory investigations and problems in water treating.

The 600 TPD H-Coal pilot plant at Cattlettsburg, Ky. will have a water treating system consisting of 2-stage stripping, API separation, emulsion holding, chrome reduction, induced air flotation, biological oxidation, clarification and biological sludge handling. No reuse of treated wastewater is contemplated, but a good opportunity to pilot reuse schemes certainly exists here. Plant completion is anticipated about the end of 1978.

The Pittsburgh Energy Research Center (PERC) has done small scale treating experiments at Bruceton, Pa., and presumably these will continue. Cooperation with vendors could prove very helpful at that location, where both Synthane and Synthoil pilot plants exist side by side.

Low Btu Gasification--

Athough to our knowledge no water treating papers have been published, it has been disclosed that Combustion Engineering has started up a 120 TPD high temperature entrained flow pilot plant in Windsor, Conn. State officials are said to be making tests on emissions and effluents. This would appear to offer a good opportunity to confirm treating methods used and treating results, and to obtain samples. DOE is funding two-thirds and C-E and EPRI one-third.

Liquefaction--
DOE has rebuilt the Conoco plant at Cresap, W.Va. Monthly reports indicate that at least sour water strippers are supplied. Other treating has not been clearly defined. Opportunities would appear to exist at this location for water treating experimentation.

EPA and Proprietary Programs

Dr. Phillip Singer at the University of North Carolina has an EPA contract for water treatment of coal conversion effluents. Publications thus far indicate that he has investigated biological oxidation and carbon adsorption, mostly on synthetic sour water mixtures or single compounds. Continuation of the experimentation is expected to yield useful information on the effects in treatment processes of specific substances encountered in coal conversion processes.

Gasification processes developed by private funds include Texaco, BGC/Lurgi and Shell/Koppers. Only BGC/Lurgi may be obligated to publish soon since they have DOE funds for design. Texaco has started up a 144 TPD plant in Oberhausen, West Germany in cooperation with Ruhrkohle A/G and Ruhrchemie. Shell/Koppers have a 150 TPD operating pilot plant in Saarland, West Germany. No details on treating have been published for either of these.

NEED FOR DEMONSTRATION OF COMMERCIAL PROCESSES

Reference has been made in previous sections to the need for demonstration of the individual commercial water treating processes and the integrated schemes proposed for coal gasification with and without p/o/t production and coal liquefaction. It has been pointed out that treating results of virtually every step should be verified by actual testing on wastewaters from the processes involved.

Some of the steps may be sufficiently verifiable on wastewaters from DOE pilot plants, but the ultimate confirmation should come from integrated operation on a demonstration plant. The demonstration plant presumably would be large enough so that operations would be directly relatable to a commercial plant in every respect.

Most of the DOE pilot plants are not directly relatable to commercial operation, since water quantities or concentrations of pollutants are frequently different due to use of once-through quench water, experimentation with steam/coal ratios and dilution with various waste streams from the pilot installation (run-off water, aqueous wastes from the laboratory, area drains, inert gas and hydrogen generators, etc.) which are much larger in relation to the feed coal than would be the case in a commercial size plant.

Some of the pilot plant equipment producing water would not be used at all in a commercial plant (e.g., hydrogen generation systems would be different and "thermal oxidizers" installed in many pilot plants in lieu of water treating would not be adequate commercially).

Commercial designs differ from the pilot installations in coal transport media (water vs. oil) and in design and operation of quenching systems. In most cases the pilot plants are designed to confirm operating conditions of gasification or liquefaction; water treating, or waste control in general, is a very secondary consideration.

Regardless of the differences cited above between pilot plant and commercial design systems, the pilot plant wastewater should certainly be usable in setting conditions for design of wastewater treating systems for demonstration plants. Small scale treating such as that being done by the Carnegie-Mellon team, the work by the Synthane group at Bruceton, Pa. and potential testing at the SRC pilot plant (as suggested in "Need for Laboratory Data on Available Schemes" - subsection "DOE Treating Programs") could be very valuable in this respect. Contract treating such as that done by AWARE, Inc. for Ashland Oil for the 600 TPD H-Coal pilot plant is a good example of another route to obtain data for water treatment system design.

It has also been urged that DOE make use of vendor and licensor know-how and facilities in the testing phases, since their equipment will be used in the commercial plants and demonstration plants. Many of these firms have good testing laboratories for obtaining the data needed to select the equipment which will be used commercially. Some of these companies also offer small rental units which could be used in the DOE pilot plants. The refining industry, for example, is quite aware of such units and have made good use of them as have other industries who are presently forced to improve their water treating systems in order to satisfy the Federal, State and local regulations that are already in force.

Demonstration of "zero discharge" water treating schemes employing recycle of treated water should be incorporated in demonstration plant designs. Specialists in cooling tower and boiler operations should be given subcontracts to participate in the design of these systems and also to monitor the systems during demonstration plant operation. In this way the best additives and conditions for control of scaling, algae, foaming, and corrosion could be established in the particular plant to be operated. Different plant locations, different raw water and coal compositions, and the different coal conversion processes, may present unique problems which could dictate additional equipment or different additives to control corrosion, scaling, algae, and foaming.

It has also been demonstrated that there are alternate integrated schemes which could be cheaper to build or operate, but which require experimental verification. Incorporation of parallel alternate water treating trains on the demonstration plants is suggested where design studies dictate that such systems are feasible if long-term operation bears out the design assumptions. Total or partial by-passes for various steps could also be installed to allow establishment of the need for that step or the minimum degree of cleanup actually required by the cooling tower or steam systems.

NEED FOR FURTHER STUDY

In addition to the further study suggested in foregoing sections ("Integrated Schemes for Wastewater Treatment," "Need for Laboratory Data on Available Schemes" and "Need for Demonstration of Commercial Processes"), study of alternate commercial technology and developing technology could prove fruitful should the schemes suggested not perform as expected.

Alternate Commercial Technology

Control technology that has been mentioned, and references documented, is as follows:

- Reverse osmosis on sludges with evaporation of reject stream. Treated water to cooling tower.
- Reverse osmosis on sludges with evaporation of reject stream. Treated water to demineralization for high pressure steam boiler feed water use.
- Study of stage-wise water condensation to concentrate inorganic materials with a minimum of organic materials.
- Establish best side-stream treatment system for cooling tower.
- Establish best system for biological oxidation. The many variations offered include trickling filters, rotating biological disc contactors, fluidized sand beds,and High Purity Oxygen Activated Sludge. Licensors for the above have been documented and could be given samples and subcontracts sufficient to establish efficiency, capital cost, and operating costs on a firm-bid basis.

- Anaerobic digestion as a first stage, followed by an aerobic second stage. One licensor of this technology has been identified and others are available.

- Powdered activated carbon addition to activated sludge systems. High surface area carbon and long sludge age, without regeneration, is one alternate. Regeneration by wet air oxidation and higher PAC rates is another. This system definitely enhances nitrogen compound removal as well as BOD and COD removal vs. conventional activated sludge with no carbon addition.

- Thermal oxidation of wastewater (Zimpro) and catalytic oxidation of wastewater should both be tried. The final cleanup step following these must be established (probably biological oxidation at second stage conditions, preferably with powdered active carbon addition). High Purity Oxygen Activated Sludge (UNOX) is another candidate for the cleanup stage.

- Establish the role, if any, of chemical oxidants such as ozone, chlorine, or hydrogen peroxide.

- Establish the role, if any, of encapsulation processes such as Chem-Fix, IU Conversion, and others.

Developing Technology

Developing control technology that has been documented includes the following:

- Biological oxidation in fluidized bed (Ecolotrol, Inc.).

- Oil fluidized evaporation (Dehydrotech Corp.).

- Heavy metal removal by SULFEX process (Permutit).

- Ultraviolet irradiation with ozone.

- Lime sludge recovery (Dorr-Oliver and others).

- Catalytic sludge precipitation (Permutit).

- Freezing (Fluor).

- Thermal incineration in reducing atmosphere and recycle to process of purge liquors from redox systems such as Stretford and Takahax (Nittetu Chemical Engineering. Ltd.).

- Coalescence of emulsified oil-water mixtures in solid beds. Substitute for API separator.

- Lindman precipitator (Precipitator, Inc.). Uses SO_2, lime and iron to remove suspended solids, BOD, oil,and grease.

- Super bacteria strains for greater cleanup in biological oxidation (Polybac Corp.).

- Mercury removal processes (Japanese "re-elixirization," FMC, Georgia-Pacific).

Licensors of the alternate and developing technology, or of technology not in wide use at present, could be furnished subcontracts to demonstrate superior efficiency or advantages in operating or capital cost. Should special problems still exist after the commercial technology schemes have been thoroughly evaluated, some of these processes may provide answers to the problems.

It is also conceivable that unusual contaminants in specific coals would not be removed by the conventional processes and could require addition of some of the alternate and developing methods to remedy this situation.

Economics of any of the methods mentioned in this section should be more thoroughly evaluated and the efficiency documented by testing on the actual wastewater from the coal conversion process in question.

The data presently being generated by DOE and EPA but not yet published could conceivably change the tentative conclusions that have been reached concerning water treating. These reports, when available, should be monitored to ascertain whether any such changes are justified.

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